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Evaluations of Novel Particulate Control Devices

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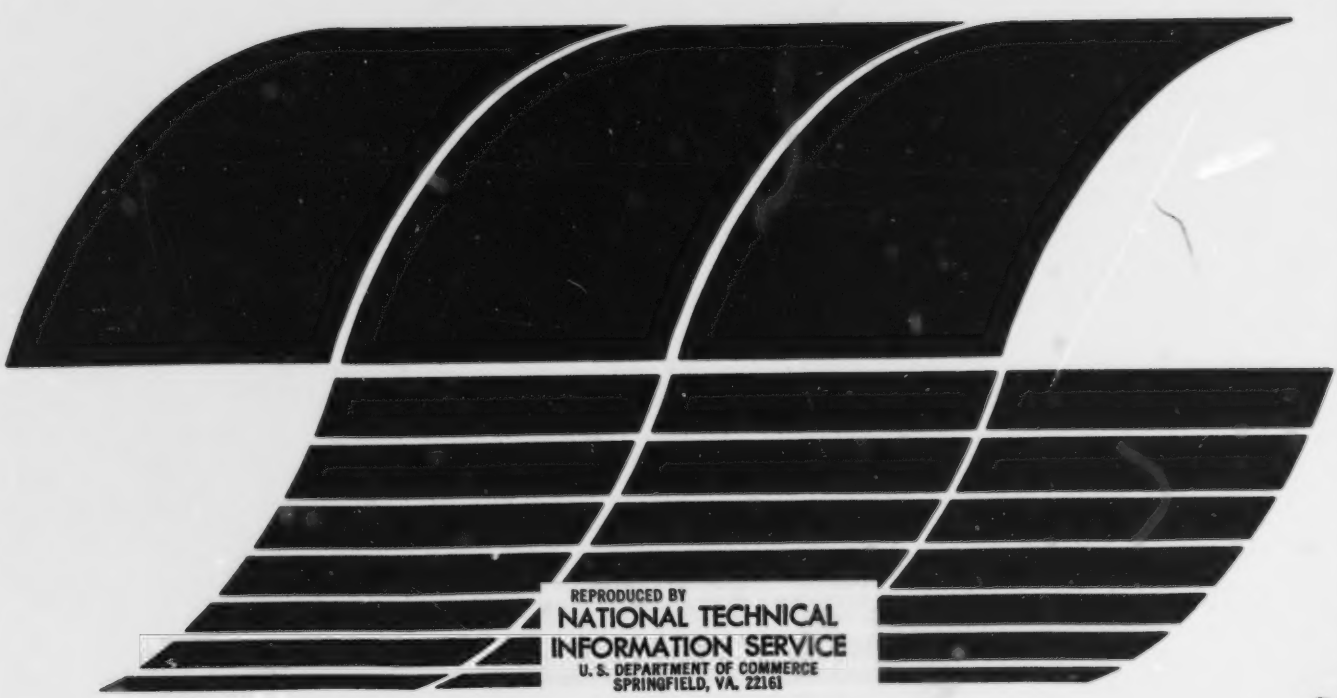

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Evaluations of Novel Particulate Control Devices

by

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16

TABLE OF CONTENTS

	Page No.
SECTION I - CONCLUSIONS	1
SECTION II - INTRODUCTION	2
SECTION III - MEASUREMENT TECHNIQUES	4
SECTION IV - ARONETICS SCRUBBER	7
Manufacturer's Description	7
Test Results	12
SECTION V - CENTRIFIELD SCRUBBER	21
Manufacturer's Description	21
Test Results	22
SECTION VI - GRAVEL BED FILTER	33
Manufacturer's Description	33
Test Results	37
SECTION VII - VARIABLE THROAT VENTURI SCRUBBER	63
Manufacturer's Description	63
Test Results	69
SECTION VIII - SUMMARY	88
BIBLIOGRAPHY	92

ABSTRACT

This report presents the results of fractional and overall mass efficiency tests of four novel particulate control devices. Three of the four devices were wet scrubbers; these were an Aronetics (Chemico) Two Phase Jet scrubber, an Entoleter Centrifield scrubber, and a CEA Variable Throat Venturi scrubber. The fourth device tested was a Rexnord Gravel Bed Filter. The devices, in the order listed, were used for controlling emissions from a submerged arc ferro-alloy furnace, an asphalt batching plant, a pulverized coal fired utility boiler, and a Portland cement clinker cooler.

Total flue gas particulate mass concentrations and emission rates were determined at the inlets and outlets of these devices by conventional techniques. Inlet and outlet emission rates as functions of particle size were determined on a mass basis using cascade impactors for sizes from about 0.5 μm to 5 μm and on a number basis for sizes smaller than 1 μm using optical, and diffusional and/or electrical mobility methods.

The text of this report includes brief descriptions of the control devices and the process on which each was utilized, the measurement methods, inlet and outlet size distributions, and overall and fractional efficiencies.

This report was submitted in fulfillment of contract number 68-02-1480 to Southern Research Institute under the sponsorship of the Environmental Protection Agency.

SECTION I

CONCLUSIONS

These evaluations were conducted as a part of a series of studies being conducted by the Industrial Environmental Research Laboratory of the Environmental Protection Agency. The studies have as their goal the identification and testing of novel devices which are capable of high efficiency collection of fine particulates. The test methods used may not have been consistent with compliance type methods, but were state-of-the-art techniques for measuring mass and fractional efficiencies. The techniques used included standard mass train, inertial, optical, diffusional, and electrical mobility methods.

Not all of the devices tested performed at 90% or greater efficiency in collecting particles smaller than 3 μm, an approximate criterion for high efficiency collection of fine particles. Three of the four devices were related to the class of control devices known as wet (spray) scrubbers. One of the four utilized waste process heat to provide most of the energy required to drive the scrubber and transport the flue gases.

In comparing the devices with each other and with more conventional gas cleaning systems consideration should be given to the efficiency needed, the efficiency which each can attain, and to the operating expenses and capital costs.

SECTION II

INTRODUCTION

This report presents synopses of the results of tests conducted by Southern Research Institute to determine the capabilities of four novel particulate control devices in controlling fine particle emissions. The four devices tested were full scale devices in commercial service. The goals of the tests were to determine overall mass efficiencies and fractional efficiencies of the devices under normal operating conditions; and, when possible, to determine what effect changes in operating conditions produced in the gas cleaning efficiencies.

The devices tested are presented in alphabetical order, by the name usually associated with them. The principal features of each device and the manufacturers are listed below.

- Aronetics Scrubber - uses waste heat to drive a two phase jet which produces a high energy spray that cleans and moves the gas. (Chemico Air Pollution Control Co., NY., NY.)
- Centrifield Scrubber - creates counter current vortex flow to improve the collection efficiency of its spray. (Entoleter, Inc.; New Haven, Conn.)
- Gravel Bed Filter - Dry Filtration using a fixed bed of fine gravel (Rexnord, Inc.; Louisville, Ken.)
- Variable Throat Venturi Scrubber - utilizes a variable cross-section venturi throat to provide control of scrubber energy usage. (Combustion Equipment Associates, Inc.; NY, NY)

The following sections provide brief descriptions of the measurement techniques used in the test programs, descriptions of collectors that were tested and summaries of the test results. More detailed information on each test can be obtained from the individual test reports which are listed in the bibliography.

SECTION III

MEASUREMENT TECHNIQUES

A total of five measurement techniques were used during the tests. These were: (1) diffusional techniques using condensation nuclei counters and diffusion batteries for determining concentration and size distribution on a number basis for particles having diameters less than approximately 0.2 μm, (2) optical techniques to determine concentrations and size distribution for particles having diameters between approximately 0.3 μm and 1.5 μm, (3) electrical methods to determine concentrations and size distribution on a number basis over the size range from 0.01 μm to 1 μm, (4) inertial techniques using cascade impactors for determining concentrations and size distributions on a mass basis for particles having diameters between approximately 0.25 μm and 5 μm, and (5) standard mass train measurements for determining total inlet and outlet mass loadings.

The mass train measurements either utilized instack filters in a manner like that of the current EPA Method 17, or were made using Method 5. When Method 5 was not used it was because of the possibility of forming particulates by condensation of vapor phase components of the gas system (for instance H_2SO_4) which would interfere with the objectives of this study.

Diffusional and electrical mobility methods were not used in conjunction on all the tests. Rather, in most instances, one or the other was used as deemed most appropriate.

The useful concentration ranges of the optical counter, electrical aerosol analyzer (Thermosystems Model 3030 Electrical Aerosol Analyzer), and the condensation nuclei counters are such

that extensive dilution of the gas streams being sampled was required. Dilution factors ranging from about 10:1 to 2000:1 were used as required by the particular circumstances. In order to insure that condensation effects were minimal, and that the particles were dry as measured, the diluent air was dried and filtered, and diffusional driers were utilized in the lines carrying the diluted samples to the various instruments.

Because only one set of optical and electrical/diffusional sizing equipment was available, it was not possible to obtain simultaneous inlet and outlet data with these methods. The system would first be installed at the inlet sampling location, and all the inlet data would be obtained. Subsequently, the equipment would be moved to the outlet and the necessary outlet data would be obtained. For the purposes of calculating the efficiency of the control device, the assumption was made that the process was sufficiently repetitive that the inlet data, as obtained above, were valid representations of those which would have been obtained during the times the outlet measurements were made. Accuracy in the measurements were often limited by process variations and the efficiencies derived from these data are somewhat uncertain. However, the trends in the fractional efficiencies derived from the data are probably real and the reported fractions of the influent material penetrating the control devices are believed to be generally correct to within a factor of two.

The optical data are presented on the basis of equivalent polystyrene latex sizes and the indicated sizes can differ from the true sizes by factors as large as two to three. Data obtained using this method were primarily intended as a means of real time monitoring of process changes, but also serve as rough checks on the data obtained with the cascade impactors.

Inertial sizing was usually accomplished using Brink Cascade Impactors for inlet measurements and Andersen or University of Washington Impactors for outlet measurements. Sampling was done at or near isokinetic rates. Errors due to deviations from

isokinetic sampling should be of little consequence for particles having aerodynamic diameters smaller than 5 μm. In those cases in which liquid droplets were present or in which the flue gas was at a temperature near a dewpoint, the impactors were heated to temperatures about 20°C above the flue gas temperature. This would tend to insure that the particles were dry when collected and sized.

Because of the wide disparities in the inlet and outlet mass loadings, complete simultaneity in inlet and outlet sampling was not possible. Outlet samples were generally of about one to eight hours duration while inlet samples were of about five to forty-five minutes duration. Because the inlet sampling could not correspond directly with the outlet sampling, an average inlet mass loading was synthesized for each size interval covered by the inlet impaction stages as appropriate for each outlet sample.

The sizes reported here for the inertial data are given in two forms, "aerodynamic" and "physical", or Stoke's, diameters. The "physical" diameters are based on measured or assumed true particle densities. If the true particle densities are lower than the value used, the sizes as given should be increased by a factor equal to the square root of ratio of the assumed density to true density. Aerodynamic diameters are diameters based on a particle density of 1 g/cm^3.

SECTION IV

ARONETICS SCRUBBER

MANUFACTURER'S DESCRIPTION

A pressurized, heated, liquid when passed through a properly designed nozzle will produce a two-phase mixture of vapor and liquid droplets that is an excellent cleaning medium. The droplets can be accelerated to extremely high velocity as a result of the expansion force created by a portion of the liquid being converted to vapor. The general configuration of this type of scrubber is shown in Figures 1a and 1b. The proper arrangement of components allows a draft to be induced which eliminates or drastically reduces fan power requirements. The two-phase jet scrubber produces water droplet velocities which vary with the temperature of the scrubbing fluid as shown in Figure 2. It is Aronetic's experience that jet velocities in the range of 1000 feet per second are quite satisfactory for particulate removal in the size range down to 0.10 microns. The velocity in the region immediately downstream of the nozzle is probably substantially supersonic since there is considerable evidence that sonic velocity in a two-phase mixture may be as low as 350 feet per second. However, Aronetic's believes that inertial impaction is the controlling mechanism in cleaning and that the existence, or absence, of shock phenomena associated with supersonic flow is not an advantage in the cleaning effectiveness. Thus, the velocity in feet per second is the controlling parameter rather than the Mach Number, or relationship of velocity to the local speed of sound. The device operates at a very low noise level. The sound frequency and level is similar to that generated by a garden hose nozzle.

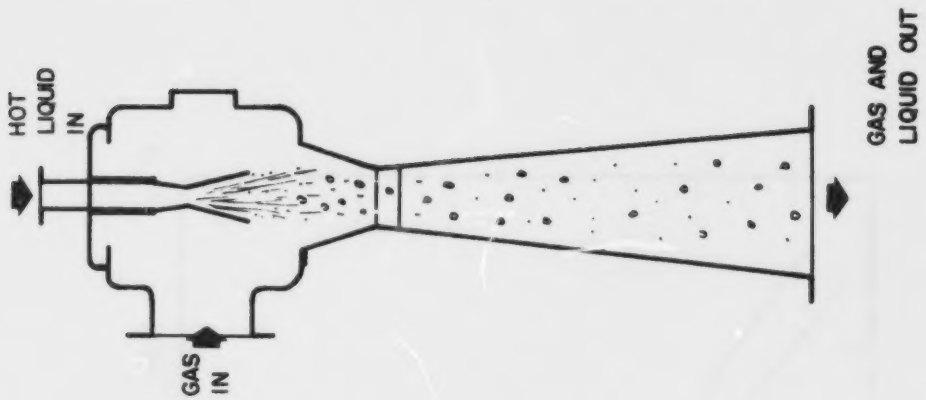


Figure 1a. Generalized Two Phase Jet Scrubber Nozzle.

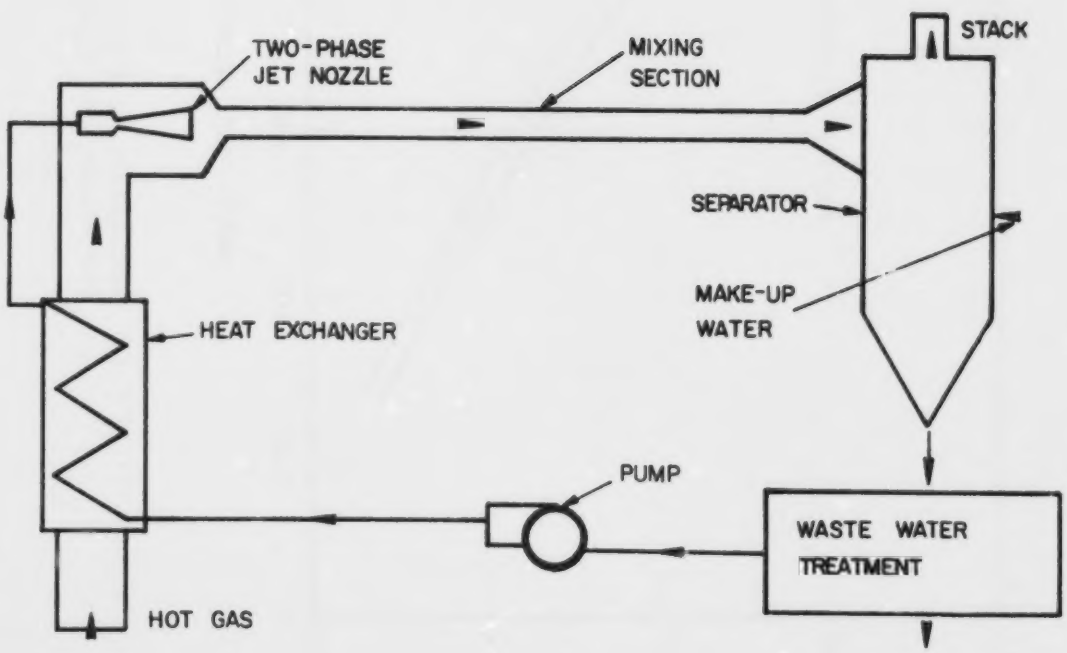


Figure 1b. Generalized Two Phase Jet Scrubber System.

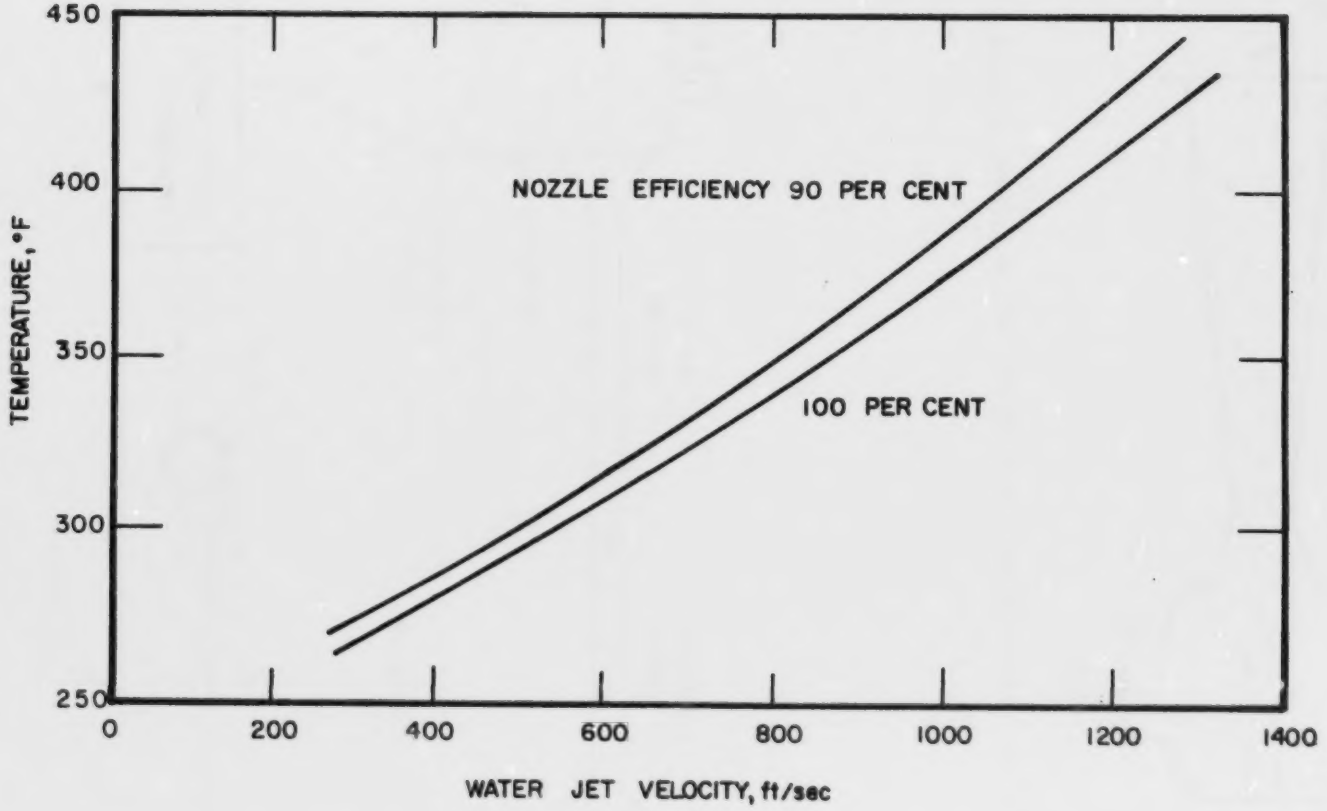


Figure 2. Variation of water droplet velocity from Two Phase Jet Nozzle with water temperature.

With respect to the draft producing capacity of this type of scrubber system, it should be evident that this is a pure momentum transfer mechanism; therefore, the amount of draft is a function of the amount of fluid passed through the nozzle and the degree to which this fluid is accelerated. The velocity is a function of initial water temperature, as shown in the previous figure. The effect of water flow rate on system pressure rise is shown in Figure 3 for water at a temperature of 400°F. It should be emphasized that this is a rise in pressure across the scrubber and should not be confused with the pressure drop which is associated with the venturi type scrubber. In most applications, the pressure rise produced by the two-phase jet is sufficient to overcome the pressure drop in other components of a complete system.

The most direct application of the two-phase scrubbing system is in the control of emissions from processes which generate high temperature gases, laden with submicron particulate. Typical examples are the various metallurgical furnaces and processes. Figure 1b shows schematically the general arrangement of the components as tested. An economizer type of heat exchanger is used to transfer thermal energy from the high temperature process exhaust gas to pressurized hot water which is delivered to the heat exchanger by a pump. Water exiting the heat exchanger is delivered directly to the nozzle in its liquid state. For most applications, the water temperature is approximately 400°F and the water pressure is approximately 350 psi, or high enough to ensure that the fluid remains in the liquid state until it has passed the nozzle throat. A properly dimensioned mixing section must be provided for intimate contact between the accelerated water droplets and the particle laden gas. The final component in the scrubbing system train is a separator which will remove the dirty water droplets and allow the clean gas to be discharged. Water drained from the separator is passed to water

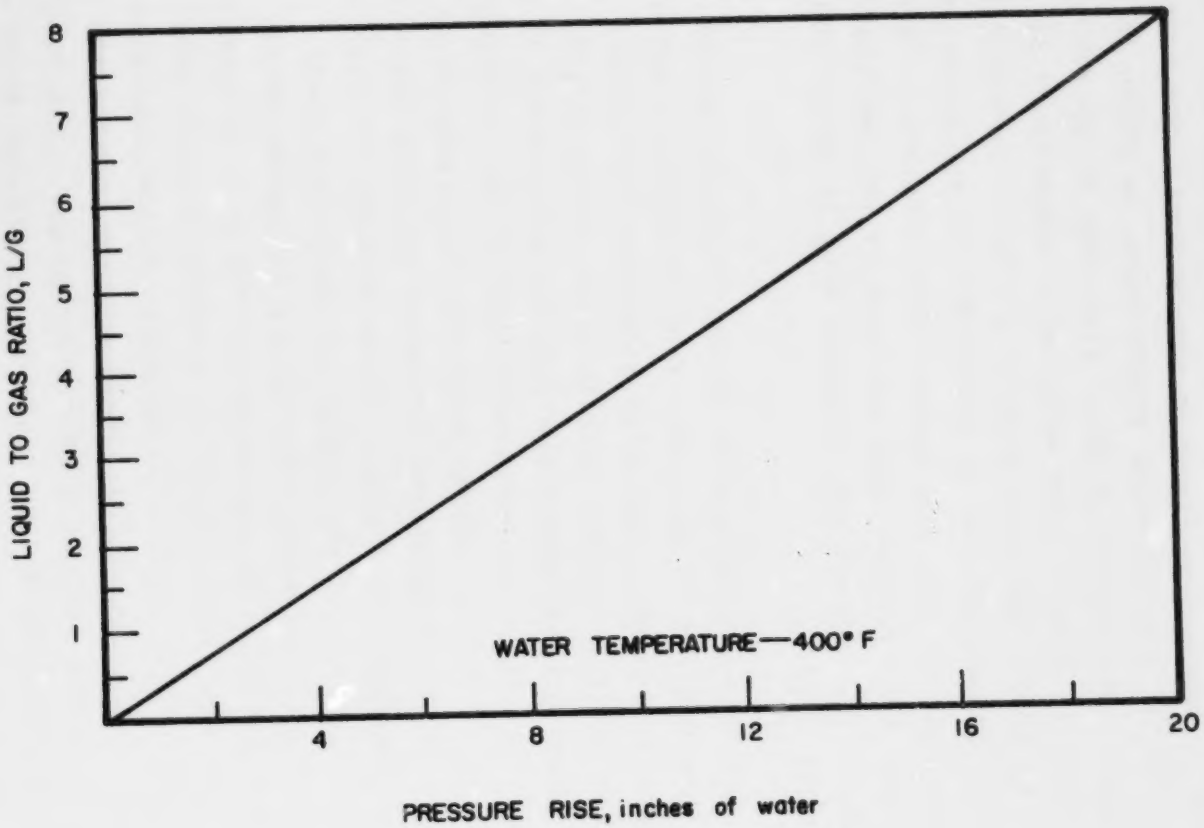


Figure 3. System pressure rise as a function of water flowrate.

treatment equipment which may be used to remove substances scrubbed from the gas and to prepare the scrubbing liquid for recycling. In the present instance, the water is used for approximately three passes through the system before final discharge.

If the process off-gas is at a temperature level above 1200°F, an additional option becomes available in the selection of components. The economizer type of heat exchanger may still be used to deliver heated water directly to the nozzle, or a steam boiler may be used as an intermediate step in the heating of water. If sufficient energy is contained in the gas, it is possible that a quantity of steam may be produced which is greater than the demands of the scrubber. This steam is then available for other possible plant applications. Other elements of the system remain essentially the same with the exception of the addition of a method to transfer thermal energy from steam to water. The choice between the steam boiler and the economizer type of system for the very high temperature gases is dictated by local conditions at the site in question.

TEST RESULTS

The tests were conducted on a 7.5 MW submerged arc ferroalloy furnace at Chromasco, Inc.'s Memphis, Tennessee facility. The furnace was operating continuously 24 hours per day, producing approximately 31 metric tons/day of ferrochrome. The furnace was tapped at approximate 2 hour intervals with several charging and stoking operatoins taking place at irregular intervals between each consecutive tap. The actual timing for any one furnace cycle varied somewhat from the nominal 120 minute tap cycle. The operations for any one batch were: (1) charging of the furnace with ore, coke, remelt, and gravel requiring about 10% of the time for one tap cycle; (2) stoking, requiring about 8% of the time over one tap cycle, and (3) tapping, requiring about 12% of the cycle. The highest particulate emission rates occur during the tapping portion of the operation, with the actual

emission rate and size distribution of the particulate being quite variable throughout the cycle. This variability caused some difficulty in both measurement and interpretation of data.

Figure 4 is a schematic of the basic furnace and scrubber system showing the inlet and outlet sampling locations. The waste process gases, at temperatures of about 920°K (1200°F) from the furnace, are carried through a duct to a heat exchanger which supplies approximately 480°K (400°F) high pressure water to drive the scrubber and provide the necessary system draft. The gas temperature leaving the heat exchanger is about 380°K (220°F). The draft for the entire furnace and scrubber system is provided by the two phase jet in the scrubber module. During operation, the scrubber produces a net pressure rise from inlet to outlet of 31.8 cm water gauge (12.5 in. H_2O) at a gas flow of 227 kg/min (500 lbs/min) and 40.6 cm water gauge (16 in. H_2O) at a gas flow of 250 kg/min (550 lbs/min). Typical volumetric gas flow rates were about 197 M^3/min (6944 SCFM) during these tests. Figure 5 shows typical averaged inlet and outlet size distributions as obtained by optical and diffusional methods over approximately one furnace cycle. Figure 8 shows the fractional efficiencies calculated from these data together with results from the impactor measurements. (In Figure 8, the impactor data are presented using particle diameters based on particle densities of 4.5 grams/cm^3.) Figure 6 shows the averaged inlet mass size distribution on a cumulative percentage basis for both aerodynamic and physical sizes. The fractional efficiencies as calculated from the impactor data are shown in Figure 7 on an aerodynamic diameter basis and in Figure 8 on a physical basis.

The results of the total particulate measurements are shown in Table 1. The overall efficiencies, by mass, based on these results are included in Table 1. Typical scrubber operating conditions are shown in Table 2.

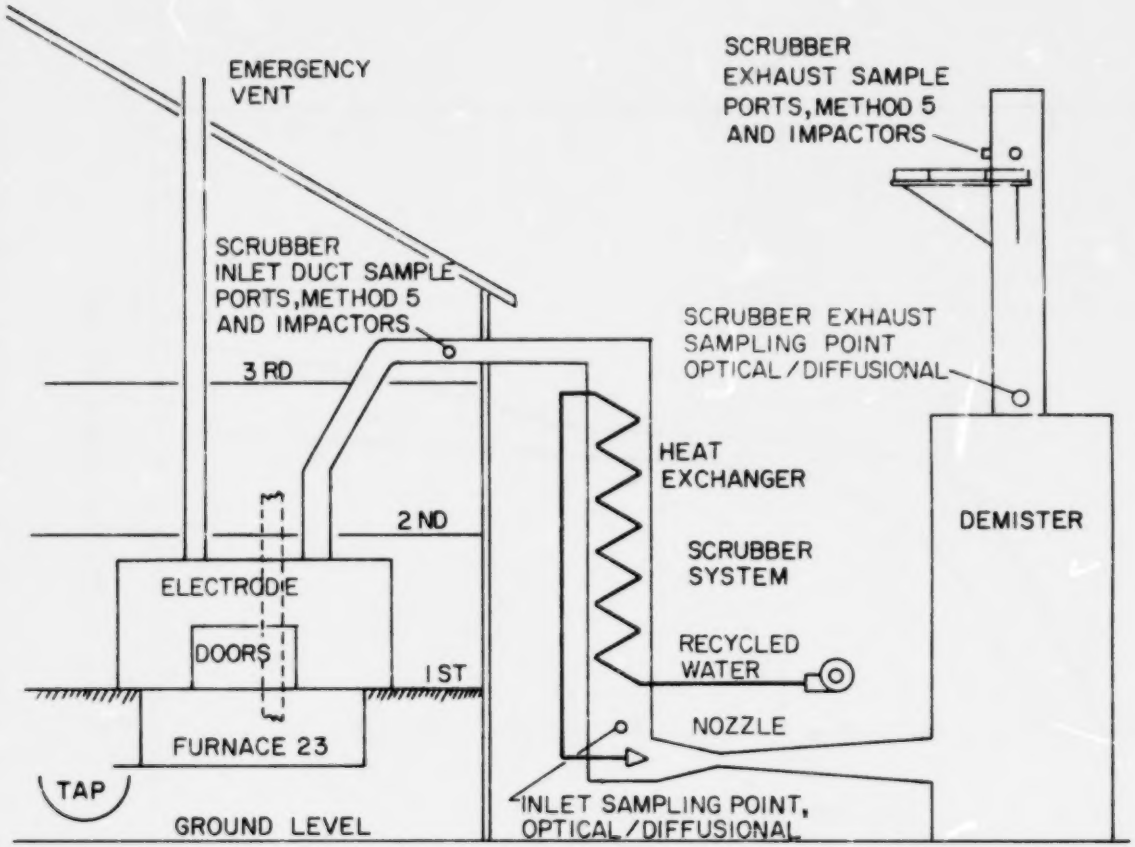


Figure 4. General layout of furnace and scrubber system showing sampling locations.

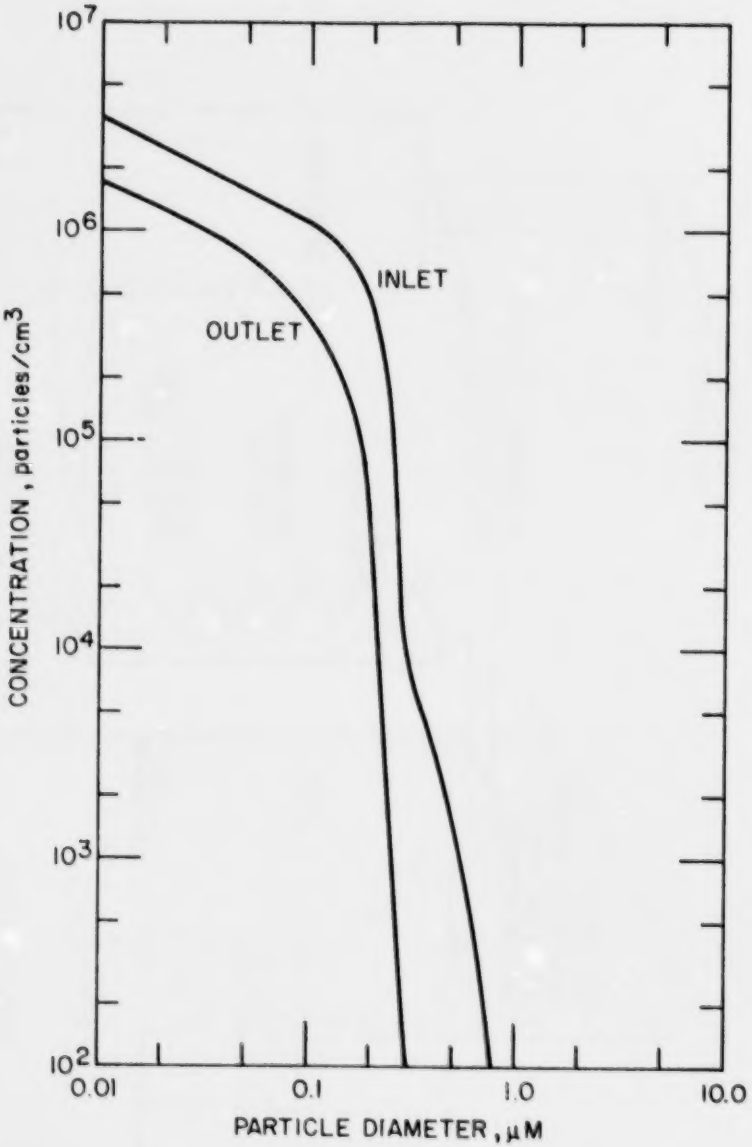


Figure 5. Average inlet and outlet size distributions on a cumulative concentration basis.

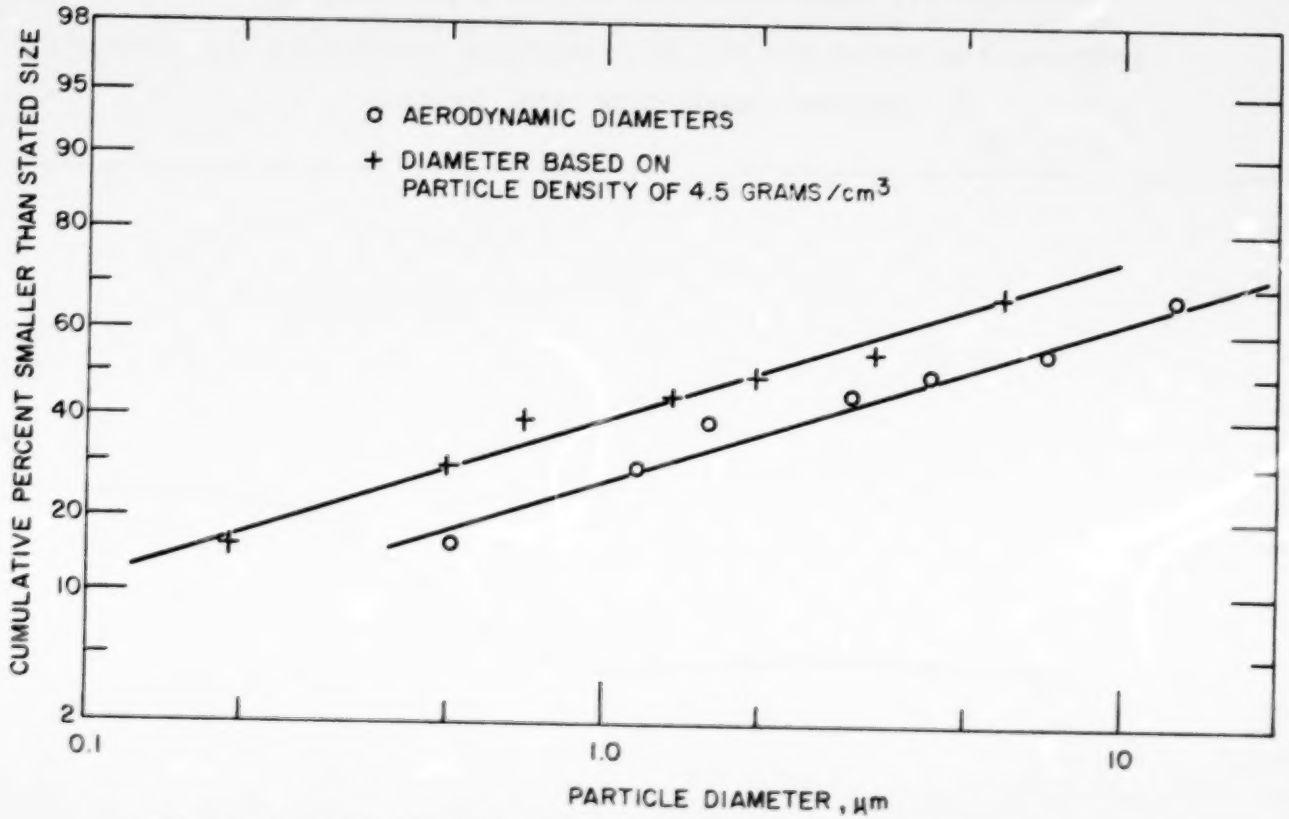


Figure 6. Average inlet size distribution on a cumulative weight percentage basis.

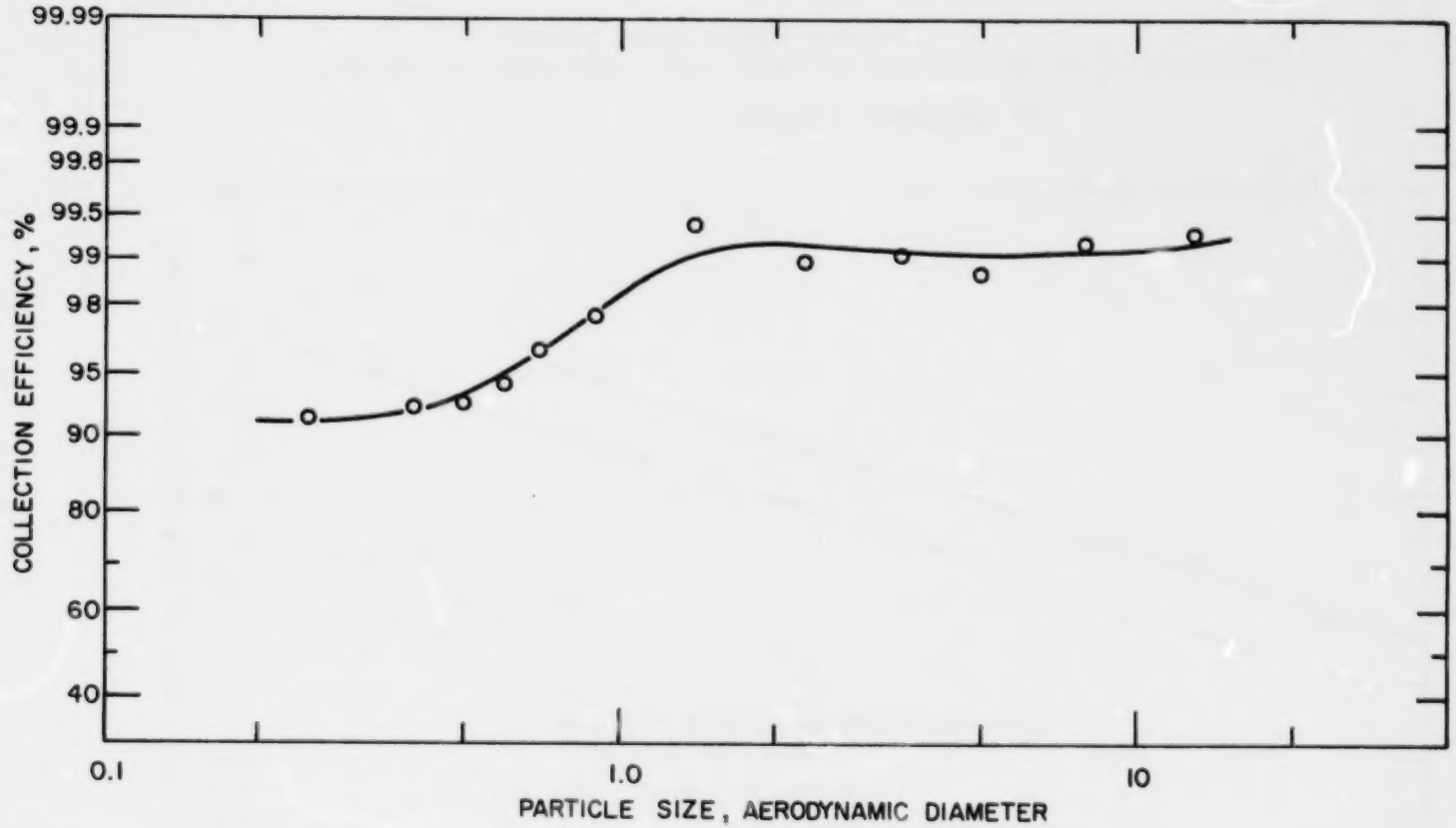


Figure 7. Fractional efficiency of the Aronetics Wet Scrubber calculated from the weighted averages of 30 inlet (Brink) and 6 outlet (Andersen Mark II) Impactor tests. Particle sizes are given as aerodynamic diameters (density = 1 gm/cm^3).

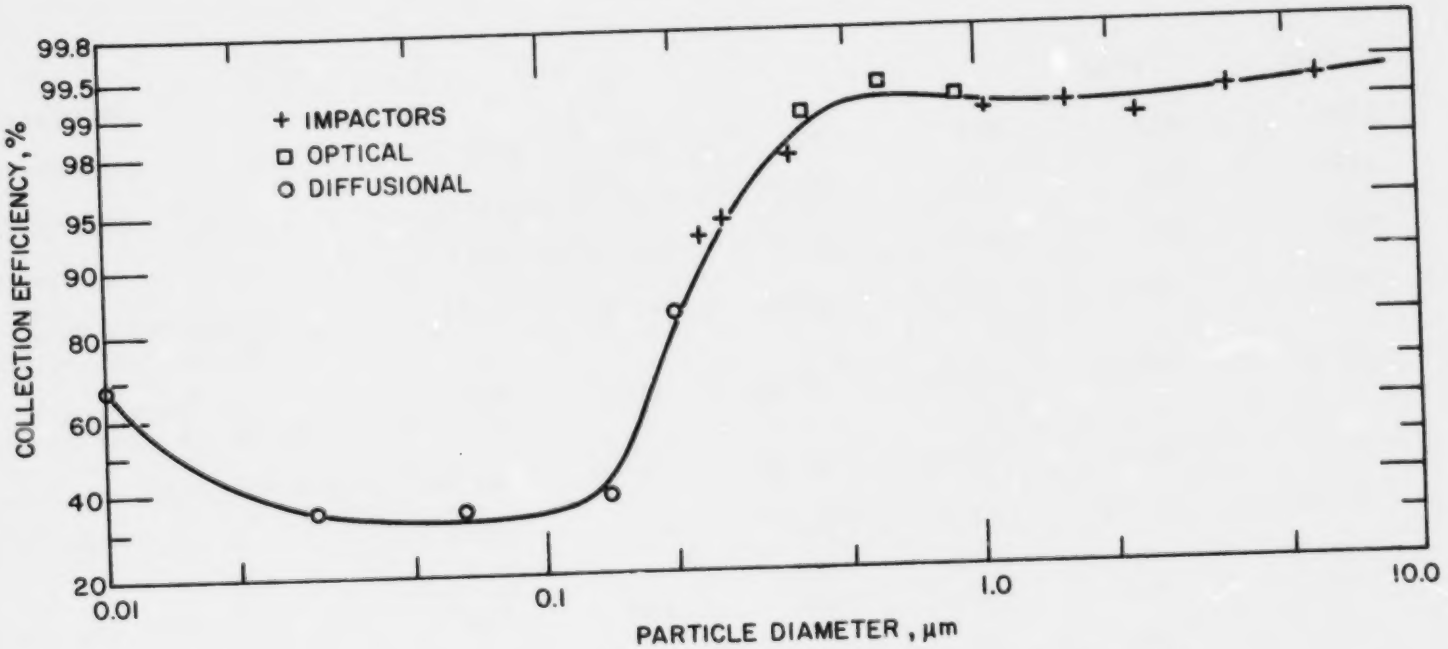


Figure 8. Fractional efficiency of the Aronetics Scrubber based on optical, diffusional, and impactor data.

TABLE 1

MASS TRAIN (METHOD 5) RESULTS

Position	Inlet	Inlet	Inlet	Inlet	Outlet	Outlet
Date	6-26-74	6-26-74	6-27-74	6-27-74	6-26-74	6-27-74
Test No.	1	2	3	4	5	6
Time Start	1045	1504	0915	1302	1047	0915
Duration, min.	180	180	180	180	360	360
Avg. Gas Temp, ^{o}K / ^{o}R	894 / 1610	894 / 1610	894 / 1610	894 / 1610	334 / 601	337 / 607
Moisture, Vol. %	4.18	3.39	5.33	3.41	22.87	20.42
Avg. Gas Velocity, M/sec	20.59	20.56	18.68	21.01	5.05	3.36
Sample Vol. at Stack Conditions, M^3	7.01	7.03	8.33	8.39	10.06	9.00
Sample Vol. at Dry Std. Conditions, M^3	2.21	2.24	2.60	2.67	6.88	6.29
Grams/ACM	.117	.181	.174	.130	.016	.011
Grams/DSCM	.366	.570	.554	.410	.023	.016
Grains/DSCF	.160	.249	.242	.179	.010	.007
Efficiency					95.1	96.7

TABLE 2. TYPICAL SCRUBBER OPERATING CONDITIONS

Water Flow	170 lpm
Water Inlet Temperature	315 °K
Water Outlet Temperature	500 °K
Air Inlet Temperature	865 °K
Air Outlet Temperature	380 °K
Air Flow	250 Kg/min.
Water Pressure	24 Atmos.

SECTION V
CENTRIFIELD SCRUBBER

MANUFACTURER'S DESCRIPTION

Modern, high-efficiency wet scrubbers for the removal of entrained particulates from industrial exhaust gases depend, almost without exception, on spray contact for particle interception. The atomization and acceleration of the scrubbing liquid into contacting zone can be performed by a wide variety of mechanical arrangements. A new high-performance contactor, the centripetal vortex contactor, is now available.

The centripetal vortex principle is illustrated in Figure 9. For any point on radius r_c and tangential gas velocity v_t, a particle may exist whose terminal velocity toward the vortex periphery exactly equals radial gas velocity v_r. If the particles consist of liquid droplets, and the peripheral inlet vanes are of the proper configuration, the vortex selectively develops a specific droplet diameter distribution within the rotating field, as determined by r_c, v_i and ϕ. The average droplet diameter created by a field of this type is substantially smaller than that created by a conventional spray contactor at the same energy level. Droplet diameter is one of the basic parameters of spray contactor efficiency.

Gas scrubber performance is most conveniently described by the relationship: $N_t = aP\gamma$, where $N_t = \log_e [1/(1-\text{collection efficiency})]$ and P = energy input per unit of gas treated. For a given dispersoid, an improvement in gas scrubber design results in an increase in a. The centripetal vortex principle should theoretically increase a by approximately 20% over a conventional contactor.

The droplet formation within a spray contactor is inherently in co-current flow in respect to the gas stream, which limits its gas absorption efficiency to a maximum of one theoretical plate per stage. Nevertheless, spray contactors, in single or multiple stage arrangements, are often the only feasible mass transfer device, especially in applications where insoluble particulate is present, or where the product of the absorption process tends to foul conventional packing and sieve plates. Mass transfer efficiency in a spray contactor is, as would be expected, a function of the L/G ratio and specific droplet surface area. Since specific surface area varies inversely with droplet diameter, it is reasonable to expect that the centripetal vortex, with its unusually small equilibrium droplet diameter, would exhibit superior mass transfer capabilities.

TEST RESULTS

The tests were conducted on a 4 ton Simplicity 200 asphalt plant equipped with a 250 ton/hr capacity H & B dryer, burning No. 2 fuel oil. The plant was operated on an intermittent basis during the test period producing approximate 300 tons of hot mix per day at rates of approximately 150 to 175 tons per hour. The plant used crushed, screened slag obtained from a local steel plant as aggregate. In the batching process, appropriate quantities of roughly sized aggregate are surface dried in the rotary kiln in order to insure that the asphalt will adhere to the aggregate (asphalt will not adhere to a moist surface). After drying, the screened aggregate is weighed in sequence and discharged from hot bins to the weigh hopper in approximately one minute. It is then discharged to the mixer, with the addition of hot asphalt requiring an additional one half minute and then immediately discharged to a truck. The complete sequence including drying requires a total time of about 30 minutes. The particulate emissions from the process are dominated by the drying, screening and mixing portions of the batch operations. Figure 10 is a schematic of the basic asphalt plant and scrubber system showing the inlet and outlet sampling locations. (The scrubber normally operates without a stack and the ducting shown beyond the scrubber exit was added specifically for these tests.)

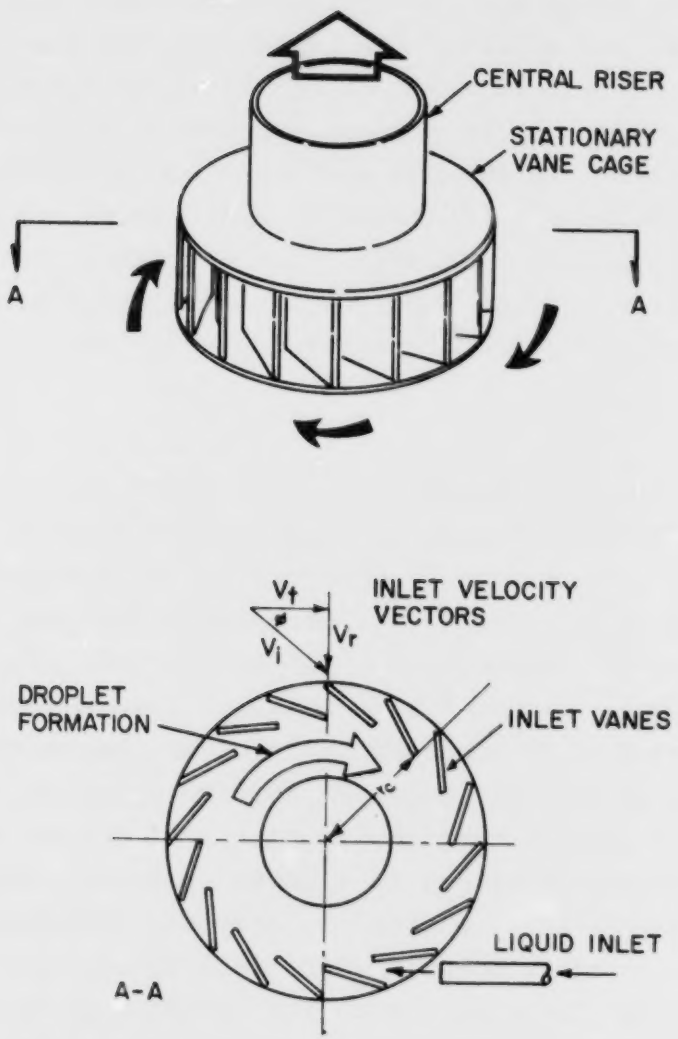


Figure 9. Centripetal vortex balances gas velocity against centrifugal force.

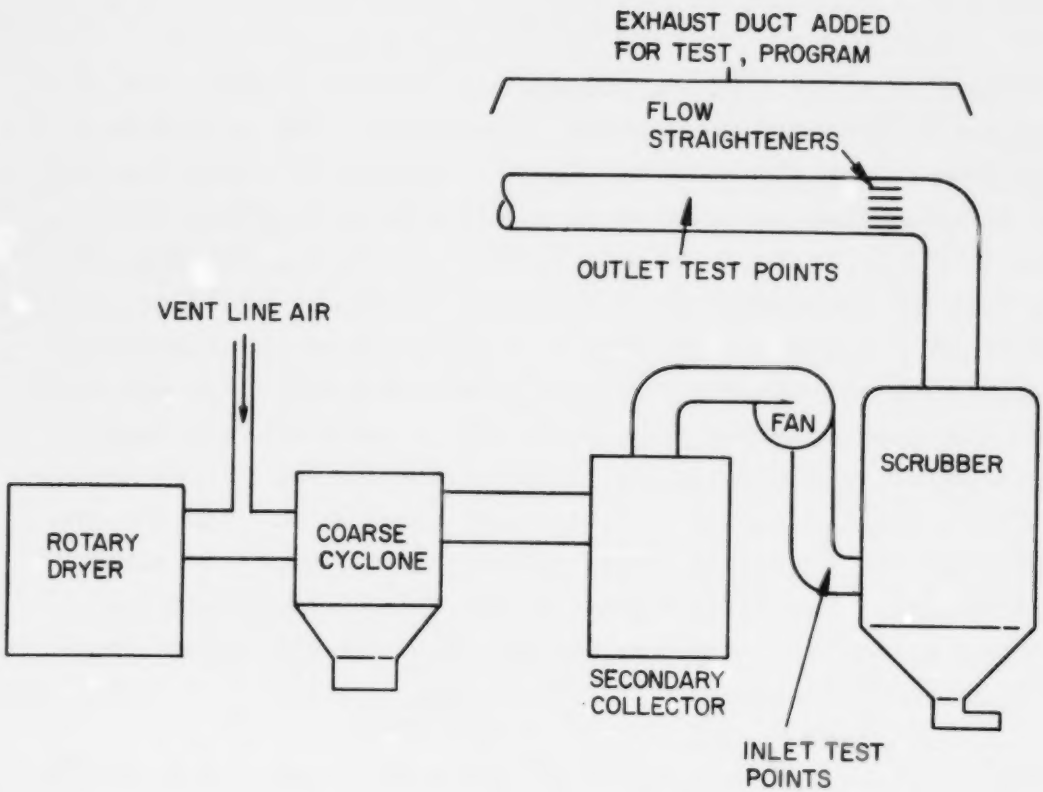


Figure 10. Schematic diagram of the asphalt plant and scrubber layout and the locations of the sampling points used in the tests.

The exhaust gases from the plant pass through a coarse cyclone to the fan at a temperature of about 55°C (130°F) and from the fan are forced through the scrubber, emerging at a temperature of about 46°C (115°F). During these tests, the scrubber pressure drop was about 17 cm w.c. (6.7" w.c.) and the system flowrate was about 11.67 M^3/sec (24,720 ACFM). The scrubber water flowrate was approximately 0.53 ℓ/M^3 (4 gal/1000 CF).

Figure 11 shows typical averaged inlet and outlet size distributions as obtained by optical, electrical, and diffusional methods over approximately one batch. Figure 12 shows the fractional efficiencies calculated from these data together with results from the impactor measurements. In Figure 12, the impactor data is presented using particle diameters based on particle densities of 2.5 g/cm^3. Because the diffusional and electrical data sets were not usually obtained simultaneously, the averages of the data sets obtained with each method were not the same. The difference was especially large at the outlet of the scrubber. Thus, the fractional efficiencies calculated from the two data sets do not precisely agree although both show the same general trends. The differences in the calculated fractional efficiencies indicate, to some extent, the uncertainties introduced by process variations.

Figure 13 shows averaged inlet and outlet mass size distributions obtained using cascade impactors on a cumulative percentage by mass basis. Figure 14 shows the same distributions on a cumulative mass concentration basis. The fractional efficiencies as calculted from these data are shown in Figure 15 on an aerodynamic diameter basis and were shown in Figure 12 on a physical basis.

The results of the total particulate measurements are shown in Table 3. The overall efficiencies, by mass, based on these results are also included in Table 3.

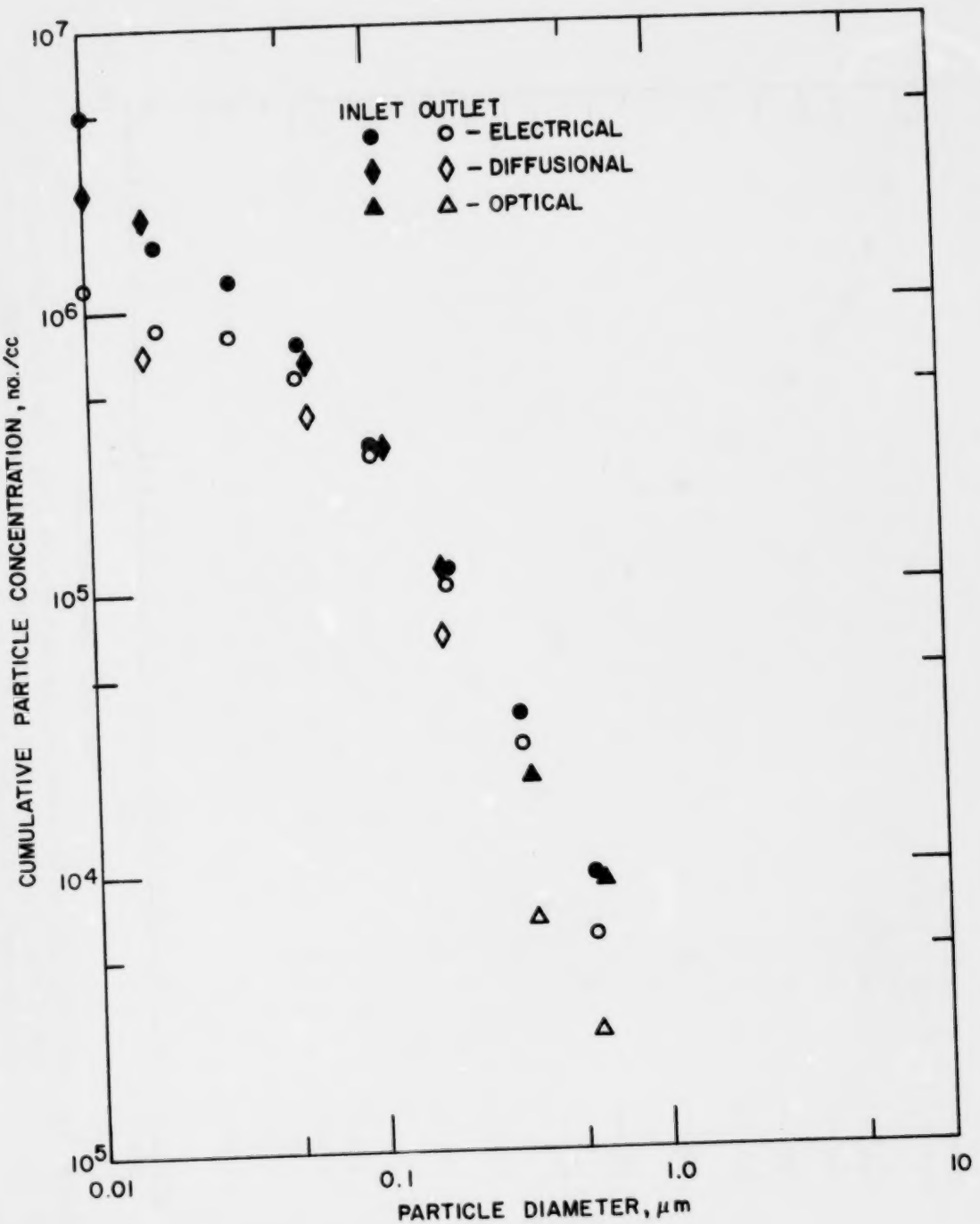


Figure 11. Inlet and outlet size distributions as obtained with optical, diffusional, and electrical techniques.

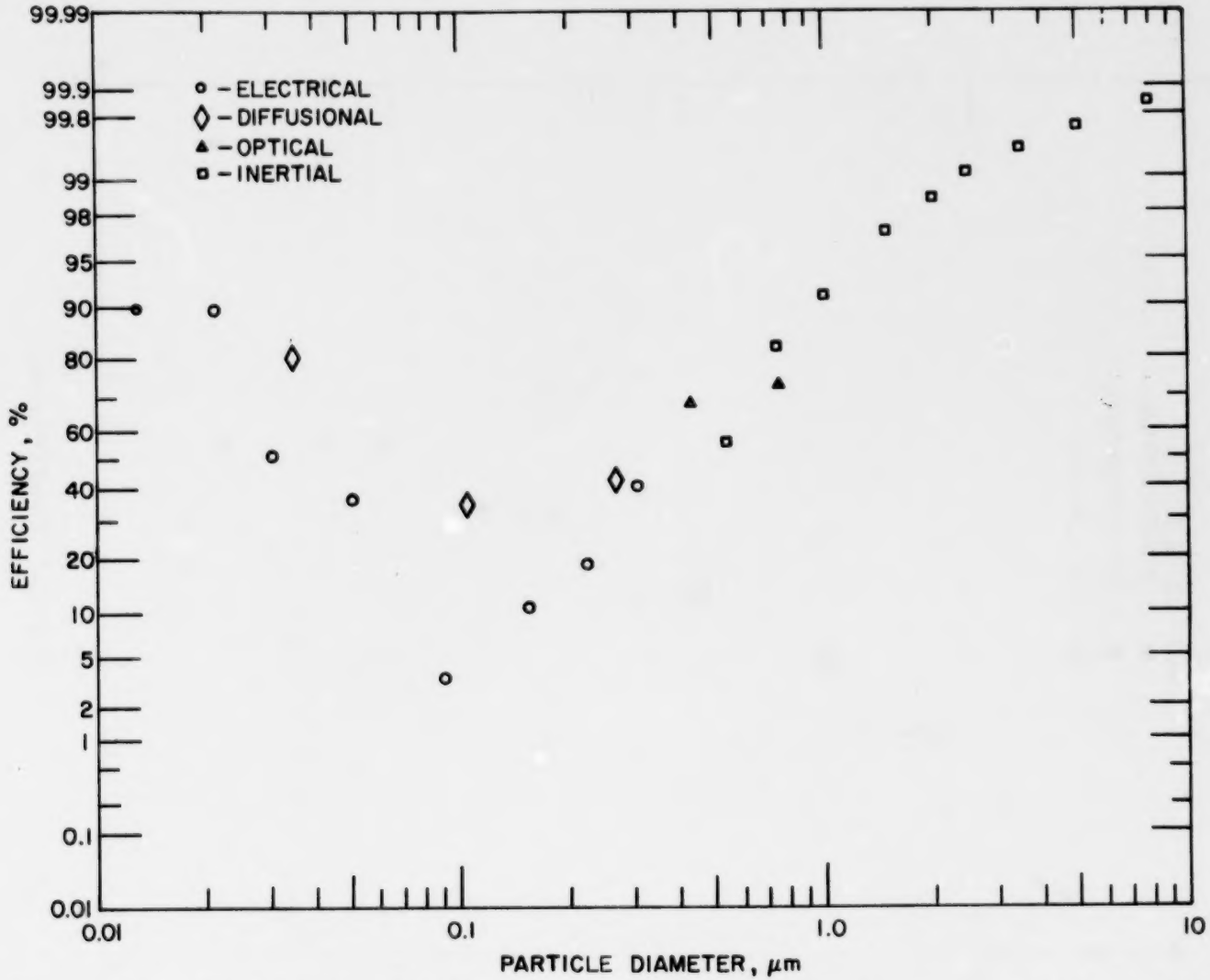


Figure 12. Fractional efficiencies as determined by the four methods used in the test program. The particle sizes shown for the impactor data are Stoke's Diameters based on a particle density of 2.5 grams/cm^3.

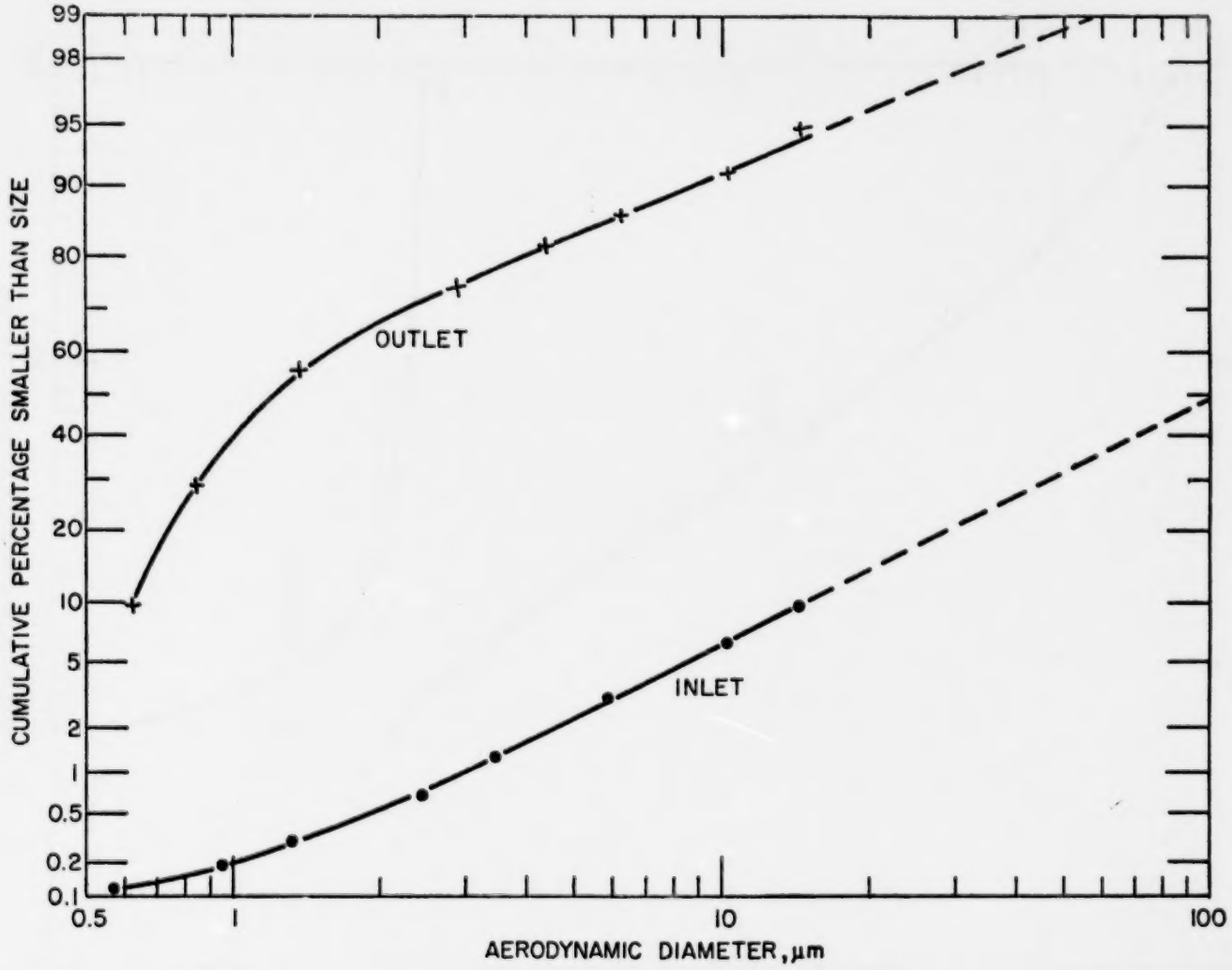


Figure 13. Entoleter Scrubber inlet and outlet size distributions on a cumulative percentage by weight basis. Data obtained with Cascade Impactors.

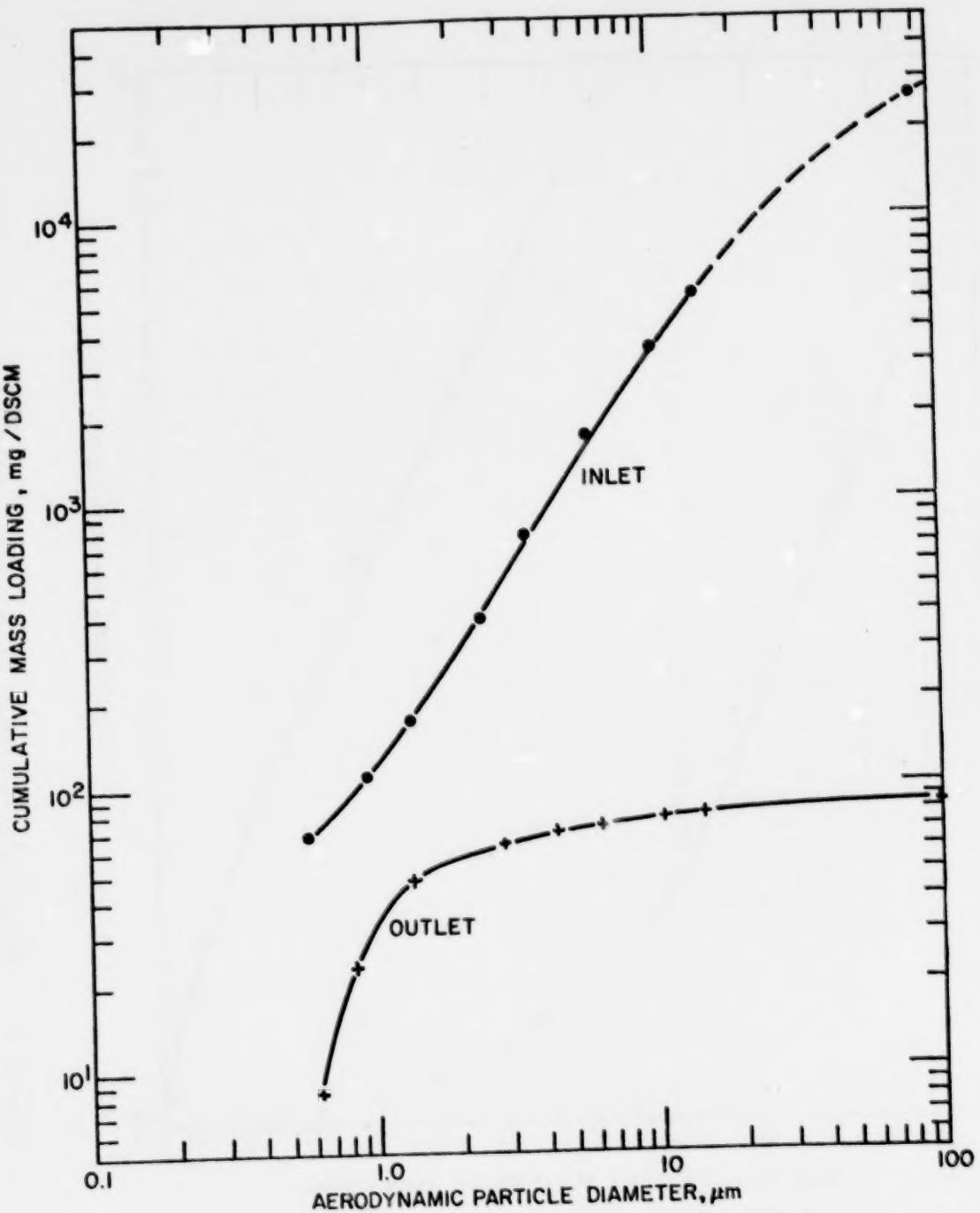


Figure 14. Entoleter Scrubber inlet and outlet size distributions on a cumulative mass loading basis.

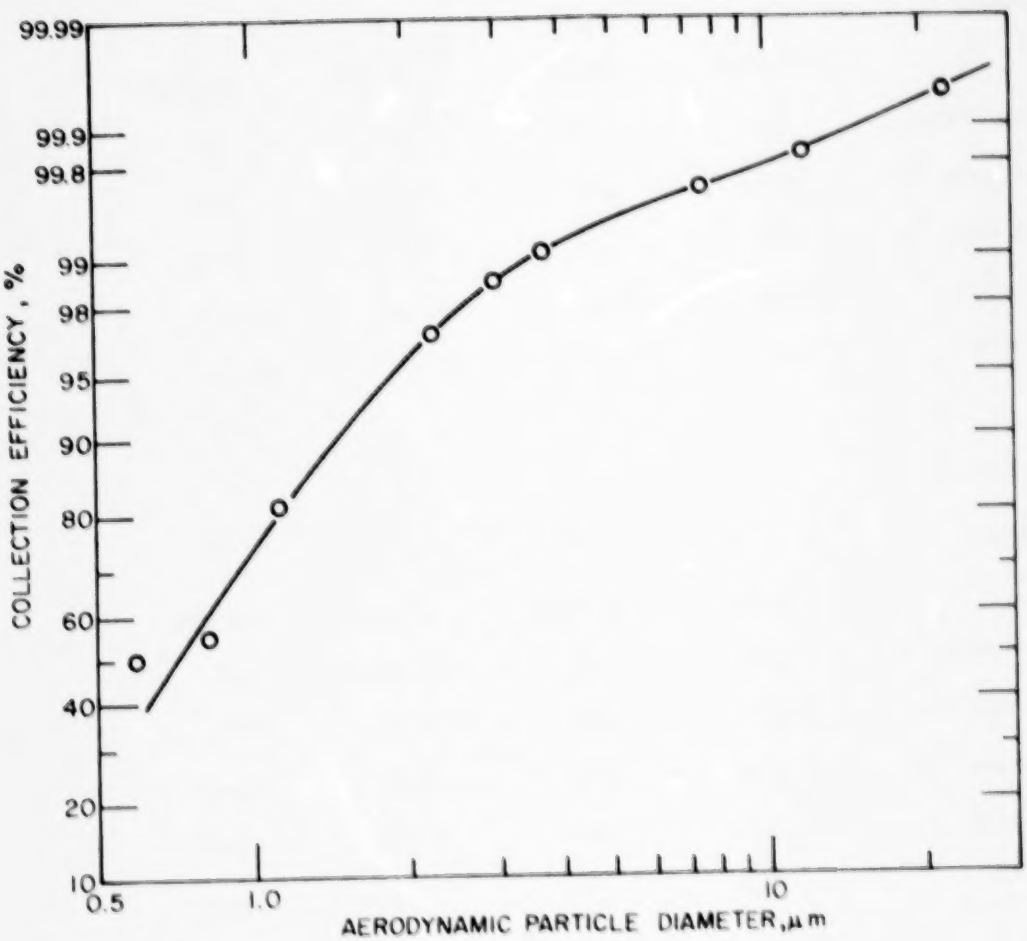


Figure 15. Fractional efficiencies of the Entoleter Centrifield Scrubber as determined from cascade impactor data.

TABLE 3

METHOD FIVE RESULTS

Inlet

Run Number	1	2	3	4
Date	10/21	10/21	10/22	10/22
Duration (Min.)	52	15	40	40
Temperature ^{o}R	602	601	582	579
Temperature ^{o}K	334	334	323	321
Grains/DSCF	8.88	25.82	14.66	28.53
Grams/DSCM	20.3	59.2	33.6	65.3

Outlet

Run Number	1	2	3	4
Date	10/21	10/21	10/22	10/22
Duration (Min.)	100	50	90	90
Temperature ^{o}R	575	583	576	574
Temperature ^{o}K	319	324	320	318
Grains/DSCF	.052	.120	.071	.045
Grams/DSCM	.119	.275	.163	.103
Collection Efficiency	99.41	99.54	99.52	99.84

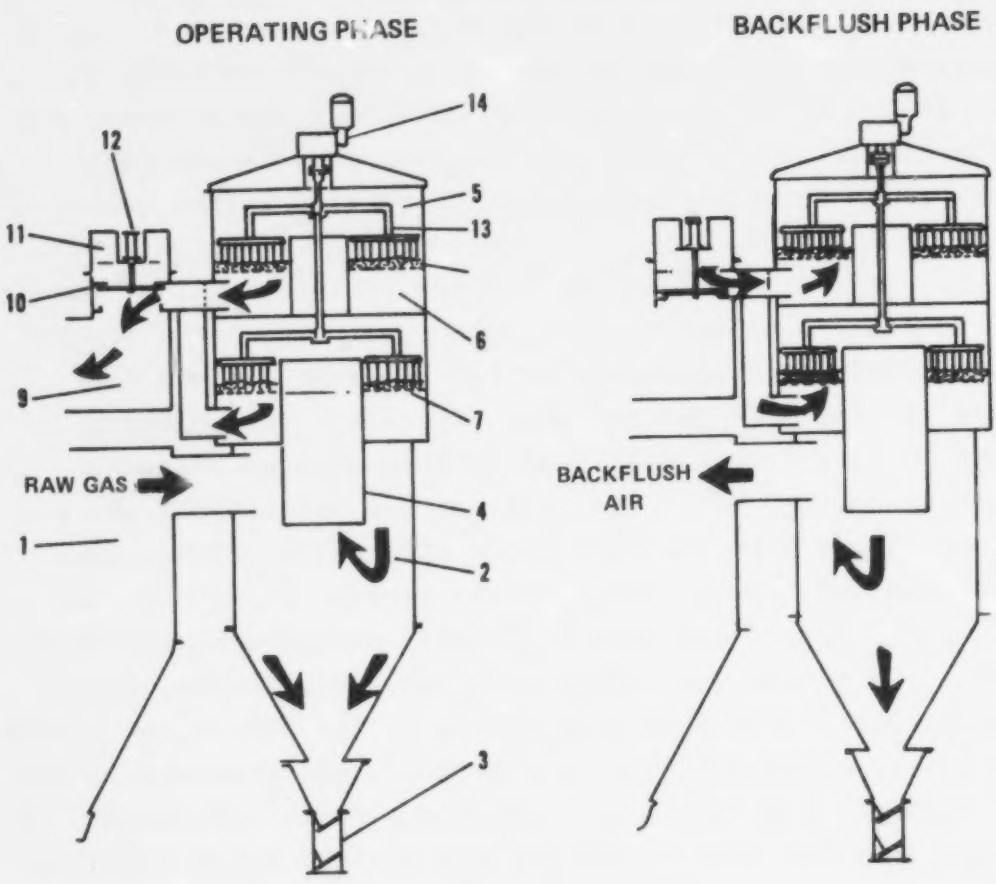
In addition to measuring particulate collection efficiencies, SO_x collection efficiencies were determined by using H_2O_2 in the impingers following the mass train to react with any SO_2 present to form H_2SO_4. Subsequent H_2SO_4 concentration determinations were made on the impinger water samples from which the SO_x collection efficiencies were calculated. These were 84%, 68%, and 88% for the three test runs for which such data were obtained.

SECTION VI
GRAVEL BED FILTE.`

MANUFACTURER'S DESCRIPTION

The gravel bed filter system is a system comprised of particulate filter beds used in conjunction with mechanical or other particulate separators in different configurations and combinations for the purpose of collecting dry particulates from dust-laden gas streams. The gravel bed filter system process consists basically of pre-separating large coarse particles from the dust-laden gas stream by means of mechanical collectors, although other means can be employed. The mechanical collector is usually the least expensive method for achieving this precleaning operation. The mechanical collectors can be in series with the filter beds, an integral part of each module, or they can be separate pieces of equipment connected in different arrangements by ductwork.

The multichamber cyclone-gravel bed filter combination consists of several units of equal size, connected by common raw gas and clean gas ducts. The operation of the system is illustrated in Figure 16. The raw gas is led into the filter via a parallel-connected raw gas plenum (1). An immediate separation of very coarse materials, by settling, takes place in this plenum chamber. From there, the gas enters the cyclone type preliminary separator (2), where the entrained medium and coarse dust is separated and removed via the discharge airlock (3) at the outlet. The precleaned raw gas now rises from the cyclone through the vortex tube (4) and enters the filter chambers (5). It passes from the top of the horizontal filter beds to the bottom, so that the remaining fine dust is deposited on the quartz grains and in the interstices of this bed (6). The cleaned gas now flows through the clean gas collection chamber (8), and passes via the 3-way



1. INLET CHAMBER
2. PRIMARY COLLECTOR (CYCLONE)
3. DOUBLE TIPPING GATE (DUST DISCHARGE)
4. VORTEX TUBE
5. FILTER CHAMBER
6. GRAVEL BED
7. SCREEN SUPPORT FOR BED
8. CLEAN GAS COLLECTION CHAMBER
9. EXHAUST PORT
10. BACKWASH CONTROL VALVE
11. BACKWASH AIR INLET
12. VALVE CYLINDER
13. STIRRING RAKE
14. STIRRING RAKE MOTOR/REDUCERS

Figure 16. Operation of gravel bed filter.

valve (10) into the clean gas duct (9). Cleaning of a filter unit can be initiated by means of a preset sequencer or by a pressure differential across the filter bed. During the cleaning cycle, the unit is isolated from the gas stream by the 3-way valve. Then, backwash air is admitted to the filter chamber in a reverse flow direction. It is either forced in by using a backwash air blower, or sucked in by negative pressure (11). The backwash air loosens the filter bed. During the cleaning process the rake-shaped double arm stirring device (13) is rotated by the geared motor (14). Thus the dust is thoroughly removed from the gravel and entrained by the backwash air, and the fluidized filter gravel is again settled after the cleaning operation. The large agglomerated dust particles are carried by the backwash air via the vortex tube (4) into the preseparator (2), where the velocity is reduced and the gas stream deflected so that a large percentage of the dust is settled out. The backwash air, containing the remaining dust, mixes with the dust-laden air in the raw gas duct and is then subjected to cleaning in the remaining units of the filter. The filter, as has been described, consists of the filter top section, containing the filter bed, and the upstream cyclone located underneath. The flow rate of the filter bed is, of course, important to the dimensioning of the cyclone, which generally has a smaller diameter than the filter section proper. Attainment of optimum flow rate for both filter bed and cyclone would result in a very large filter bed relative to the cyclone.

To efficiently use the space required for the overall unit, it is desirable to increase the bed area without increasing the bed diameter to a value much greater than that of the cyclone. As can be seen from Figure C2, this objective was obtained by placing two parallel-connected filter beds one on top of the other.

The cyclone-gravel bed filter combination is automatically controlled via a program transmitter. Following the signals of this continuously operating program transmitter, the clean gas port for a chamber to be cleaned is shut and the backwash air port

opened. The drive of the raking device is started after a time lag. The adjustment of the duration of the cleaning process and of the intervals presents no difficulty. When the cleaning process is over, the raking device is stopped first. Then, after another time lag, the clean gas port is opened; and, at the same time, the backwash air flow is stopped so that filtering can be resumed in this chamber.

The design of the double bed filter element and its stirring (leveling) device and drive is also simple.

The rake-shaped double arm of the raking device is firmly connected to the vertically arranged geared motor via a rigid coupling. This coupling is provided with an asbestos washer serving as a thermal insulation. The bearing of the geared motor is reinforced so that the forces, which result when the rake is rotating, can be absorbed without any further intermediate bearing. The filter bed is spread on a woven supporting screen made of spring or stainless steel. Thanks to the special method applied for dust removal, this supporting screen is not subject to significant wear.

As has already been mentioned, the gases enter the cyclone at relatively low velocities, so that generally no significant wear occurs in the cylindrical section. The outlet cone of the preseparator is protected against wear by means of a ceramic lining. A weighted double-acting valve assembly (tipping valves) is arranged beneath the outlet cones, where it serves as a dust discharge device and as an air seal. In many cases, this valve assembly is mechanically actuated. All moving parts required for the cleaning process will be in operation during that period only. Normally, the duration of the cleaning phase is about three to five minutes.

Once the large dense particulates have been removed in the first pass of the raw gas through the precleaning stages of the system, the gas containing the residual fines passes through the

packed solids filter beds. In this passage, the impingement of the small particulates on the discrete filter media causes an agglomeration to take place in which the fine particles are joined with other fine particles to form larger agglomerates. Periodically that particular filter is taken out of the system and back-washed by reversing a gas flow upwardly through the bed while it is being mechanically stirred by a rake. The bed is both mechanically and pneumatically disturbed and this results in a semifluidized state in which the agglomerates (which are lighter than the filter media and have a greater surface area/weight ratio) are blown backwards from the bed and are then collected in the common precleaning cyclones or recollected in the remaining gravel bed.

TEST RESULTS:

The tests were conducted on a gravel bed filter system controlling the emissions from clinker coolers on two 500-tons/day Portland cement kilns. Both kilns were operating at full capacity during the majority of the tests. In the process a slurry of water and powdered raw materials is introduced to a kiln in which they are calcined, forming clinker which drops onto a moving bed clinker cooler. A series of fans beneath the moving bed blow cool, ambient air through it to reduce the temperature of the product. The bulk of the air from the clinker cooler is used as combustion and secondary air for the kiln. Scavenge air from the system (that air not required by the kiln) is drawn off through the gravel bed filter. Because the air to the gravel bed is scavenge air, the airflow through the filter is subject to considerable variation in both temperature and volumetric flow rate. Inlet gas temperatures ranged from 77°C (170°F) to 230°C (450°F). Outlet temperatures ranged from 71°C (160°F) to 129°C (265°F). The temperature drop between the inlet and outlet results primarily from the addition of approximately 283 M^3/min (100,000 CFM) of ambient air which is used to periodically clean the gravel beds. During these tests, the gravel bed system pressure drop was approximately 12 cm water column (4.7 in). The system flow

rate ranged from about 600 to 1000 DNM^3/min at the inlet. (The latter figures exclude the backflush air). Testing took place during the months of August and November 1975. The second series of tests were done as a result of a determination by the manufacturer that the system was not operating under optimal conditions during the August test. The results of the November tests did show substantial improvement in system performance as compared with the August data.

Figure 17 is a schematic of the basic Portland cement kiln, clinker cooler, and gravel bed systems showing the inlet and outlet sampling locations.

The gravel bed filter is constructed on a modular basis, with eight modules making up the system in this instance. Each module consists of a cyclonic inlet section followed by two gravel beds operating in parallel.

At any one time, during normal operations, seven (7) of the eight (8) modules are on line in the forward flow direction with one being cleaned and renewed by backflushing with heated ambient air. In the installation tested, the modules were cleaned in sequence with a backflush time of 6 to 20 minutes. During the first 45 seconds of the backflush period, a mechanical raking system is actuated to stir up the dirty gravel. The particulate laden air from the module being backflushed is exhausted into the inlet plenum of the remaining modules. This backflush and raking operation and the frequency with which it takes place has a very pronounced influence on the collection efficiency achieved by the system as will be discussed later in this report. The tests of this unit were done under long term steady-state conditions with backwash periods of 12 and 6 minutes. In addition, limited short term tests under transient conditions were made using 6 and 20 minute backwash intervals during a period when the system was normally being operated with a 12-minute backwash interval. Design values for the system tested are given in Table 4.

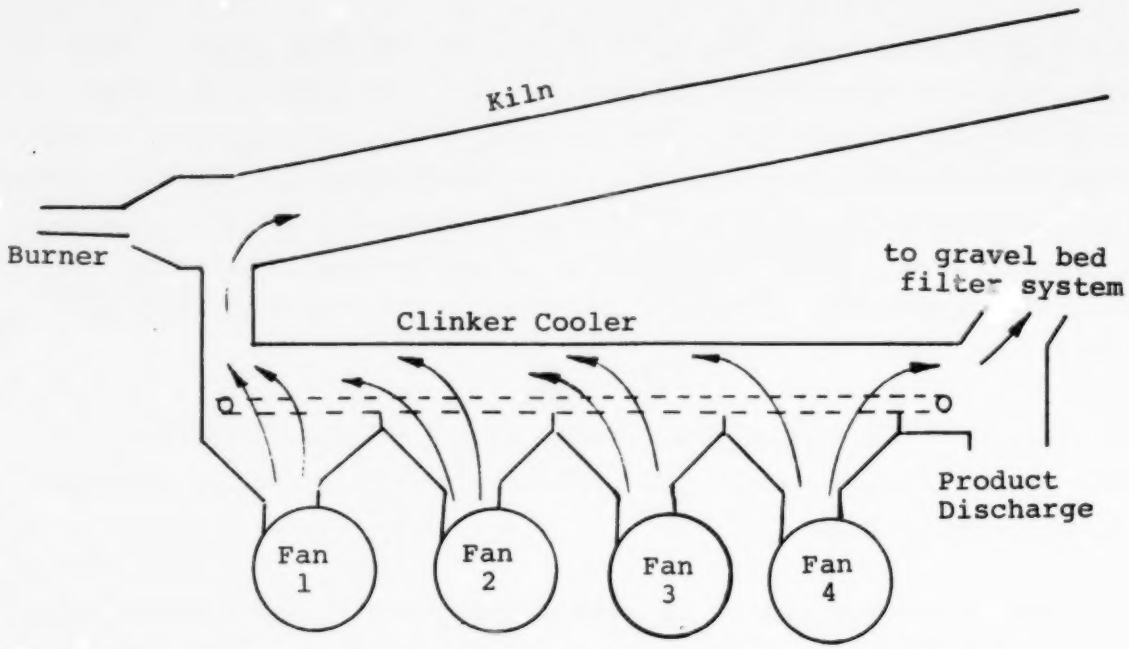


Figure 17a. Portland cement kiln and clinker cooler layout. Arrows indicate directions of air flow.

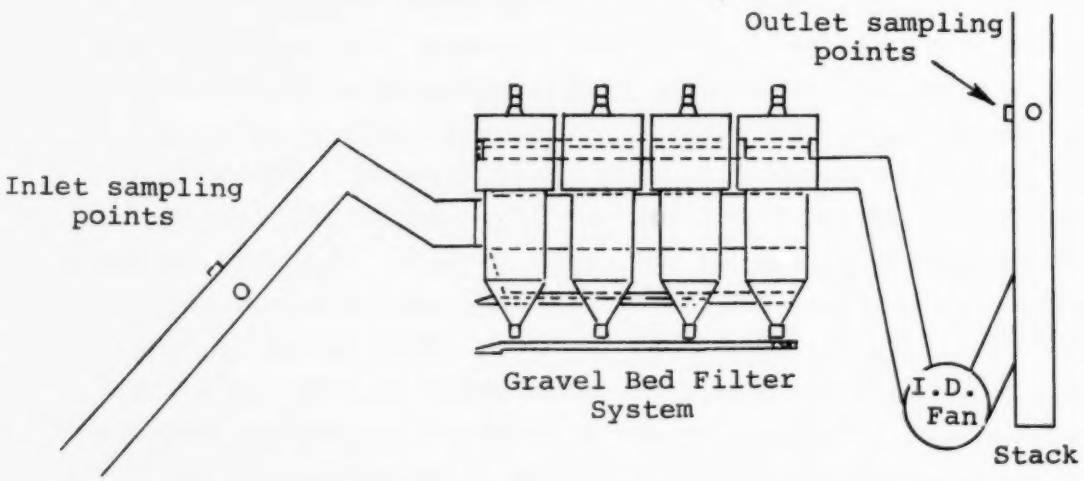


Figure 17b. Gravel bed filter system layout and sampling locations.

Table 4

Design Specifications Of The System As Tested

Inlet Volume Flow: 2266 ACM/min at 204°C
(80,000 ACFM at 400°F)

Backflush Volume Flow: 317 ACM/min at 66°C
(11,200 ACFM at 150°F)

Pressure Drop: 25.3 cm w.c. (10 inches w.c.)

Gravel Size: 4 mm (5/32 inch) x No. 6 mesh

Bed Depth: 11.4 cm (4½ inch)

Bed Area: 3.72 m^2/Bed (40 ft^2/Bed)

(For a total of 59.5 m^2 of bed area with 52 m^2 actively filtering in normal operation.)

The data obtained by Method 5 technique is summarized in Tables 5 and 6. The overall collection efficiency for each of these tests is also given in Table 6, together with the gas flow per module (at flue conditions). It would appear from these data that the collection efficiency of the system is sensitive to the gas flow per module and improved markedly with decreasing gas flow over the range of values that were obtained during this series of tests. The differences in the inlet and outlet gas flows in Tables 5 and 6 are due to the addition of the backflush air. During these tests, the backwash duration was 12 minutes.

On a service-inspection trip by Rexnord personnel subsequent to this test series, it was found that the rakes were being activated a fraction of a second to a few seconds before the backwash valves were actuated, which would tend to reduce the average collection efficiency of the device. In addition, accumulated experience with the gravel bed system in this application indicated that the beds were not being adequately cleaned with the 84 minute forward-flow/12 minutes backwash cycle, which results in a buildup of dust within the beds. This increases the energy requirements and decreases the collection efficiency of the system. Resetting the rake timers and changing the backwash intervals to provide a 42-minute forward flow/6 minute backwash cycle resulted in a substantial improvement in performance as reported by plant personnel. Consequently, a second series of tests were run during the month of November, 1975. Method 5 measurements were not made during the retests. The outlet impactor data during the second test series indicated a reduction in the outlet particulate loading by approximately a factor of 3.5 as compared to the impactor outlet data obtained during the first test series, while inlet data from cascade impactors indicated no discernable differences in loading and size distribution between the two test series. Thus, in steady state operation under a more nearly optimum operating cycle, the overall efficiency in this application is estimated to be approximately 99.5%.

Table 5

Mass Emission Tests - Method 5

Inlet

Run #	1	2	3	4	5
Date	8-25-75	8-26-75	8-26-75	8-27-75	8-27-75
Time	1350	1015	1435	1050	1515
%, Moisture	2.33	1.65	2.60	1.54	1.80
Velocity, m/s (f/s)	10.44(34.24)	9.70(31.48)	7.42(24.35)	8.50(27.89)	7.97(26.15)
ACM/min (ACFM)	1467.3(51812)	1364.5(48140)	1043.5(36847)	1195.2(42203)	1120.6(39570)
SDCM/min (SDCFM)	937.1(33089)	1088.4(38431)	761.6(26891)	875.3(30906)	739.9(26128)
Grams/ACM (Grains/ACF)	2.039(0.891)	1.144(0.500)	1.602(0.700)	2.130(0.931)	2.078(0.908)
Grams/SDCM (Grains/SDCF)	3.192(1.395)	1.435(0.627)	2.197(0.960)	2.911(1.272)	3.146(1.375
Kg/hr. (Lbs/hr.)	179.46(395.65)	93.68(206.54)	100.37(221.27)	152.84(336.96)	139.68(307.94)

Table 6

Mass Emission Tests - Method 5

Outlet

Run #	1	2	3	4	5
Date	8-25-75	8-26-75	8-26-75	8-27-75	8-27-75
Time	1400	1015	1445	1100	1515
Velocity, m/s(f/s)	8.82(28.94)	8.76(28.73)	6.79(22.29)	7.98(26.18)	7.26(23.81)
%, Moisture	2.29	1.83	1.86	1.64	1.38
ACM/min(ACFM)	1631.7(57619)	1619.9(57201)	1256.8(44379)	1476.2(52124)	1342.5(47405)
SDCM/min(SCDFM)	1239.3(43759)	1326.4(46837)	1017.5(35927)	1174.2(41461)	1049.7(37067)
Grams/ACM(Grains/ACF)	0.094(0.041)	0.030(0.013)	0.064(0.028)	0.043(0.019)	0.034(0.015)
Grams/SDCM(Grains/SDCF)	0.121(0.053)	0.037(0.016)	0.080(0.035)	0.055(0.024)	0.043(0.019)
Kg/hr.(Lbs/hr.)	9.02(19.88)	2.91(6.42)	4.89(10.78)	3.87(8.53)	2.74(6.04)
No. of Active Modules	7	7	4	7	7
Average Flow per Active Module in ACM/min(ACFM)*	233.1(8230)	231.4(8170)	314.2(11095)	210.8(7445)	191.7(6770)
Efficiency	95.00	96.9	95.1	97.5	98.0

*Includes backflush air.

The impactor data are summarized in Figures 18 and 19. Figure 18 shows averaged inlet and outlet mass size distributions on a cumulative percentage versus aerodynamic diameter basis. Figure 19 shows the same size distributions on a cumulative mass concentration basis. Inlet sampling times of three hours produced catches of substantially less than one milligram on all stages collecting particles smaller than 3 μm with the Brink impactors. These catches were considered too low to give reliable loadings so only the Andersen inlet data were used for fractional efficiency calculations.

The high concentrations of large particles in the gas streams coupled with the non-ideal particle separation and collection characteristics of the impactors tend to make the impactor filter catches difficult to interpret. In sampling particulate having the properties and size distribution of the type encountered here the impactor back-up filter catches can be dominated by oversized particles, a small fraction of which are not retained by the stages which should collect them. As a result, no fractional efficiencies were based on the impactor back-up filters and the size distributions obtained with the impactors as shown in Figures 18 and 19 are given both with and without the back-up filter catches.

The fractional efficiencies as calculated from the impactor data are given in Table 7 together with the gravel bed operating pressure loss and are shown in Figures 20 and 21. Also shown in Figures 20 and 21 are fractional efficiency curves derived from overall averages of the respective data obtained during the two test series. Because the outlet loadings appear to be decoupled from the inlet insofar as short term behavior is concerned, these overall averages may better illustrate the system performance than do the individual tests. The negative efficiencies shown for some particle sizes are discussed later.

Data on the concentration and size distribution of ultrafine particulates were determined during the first test series

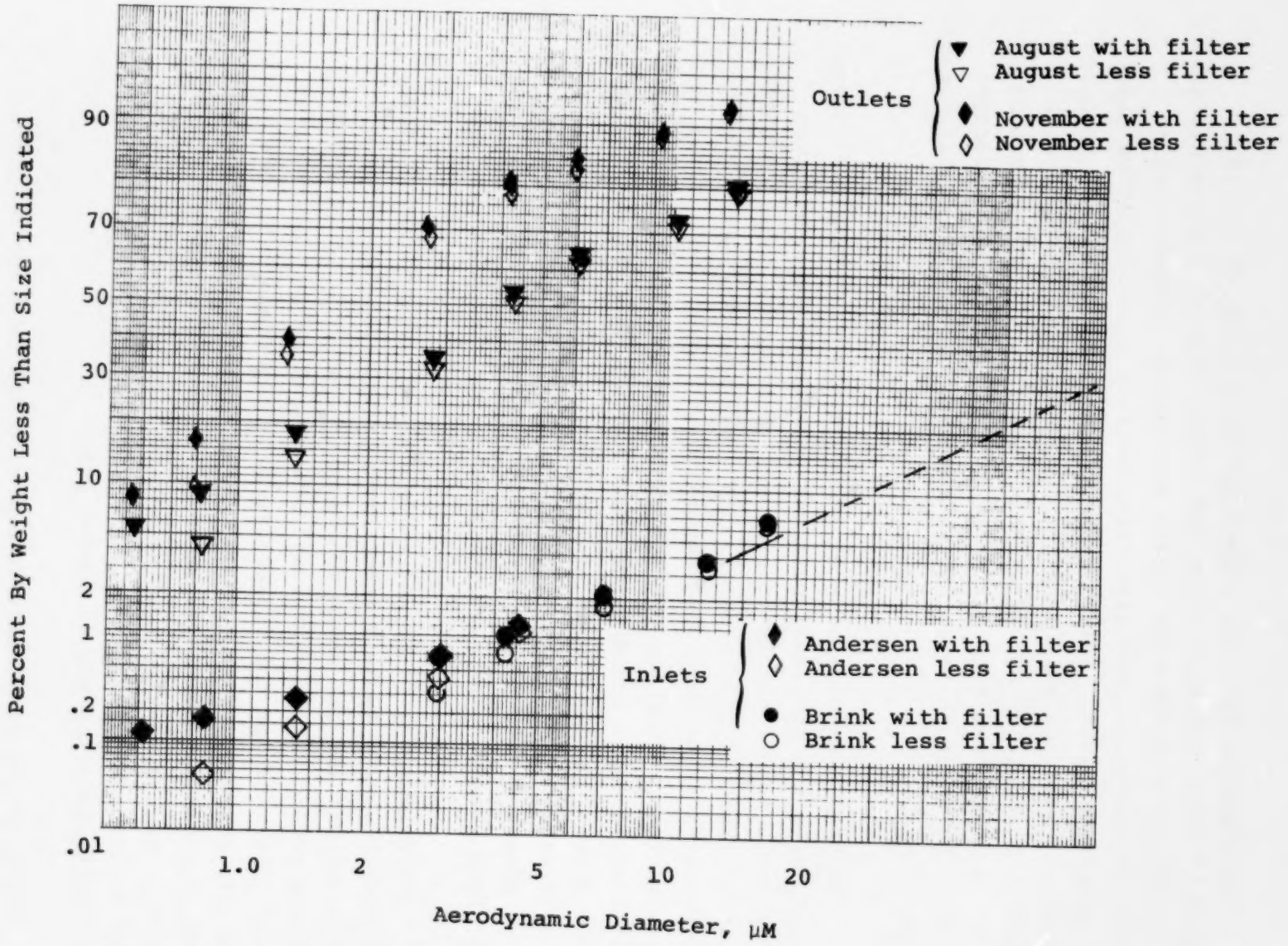


Figure 18. Inlet and outlet size distributions on a cumulative percentage by weight basis.

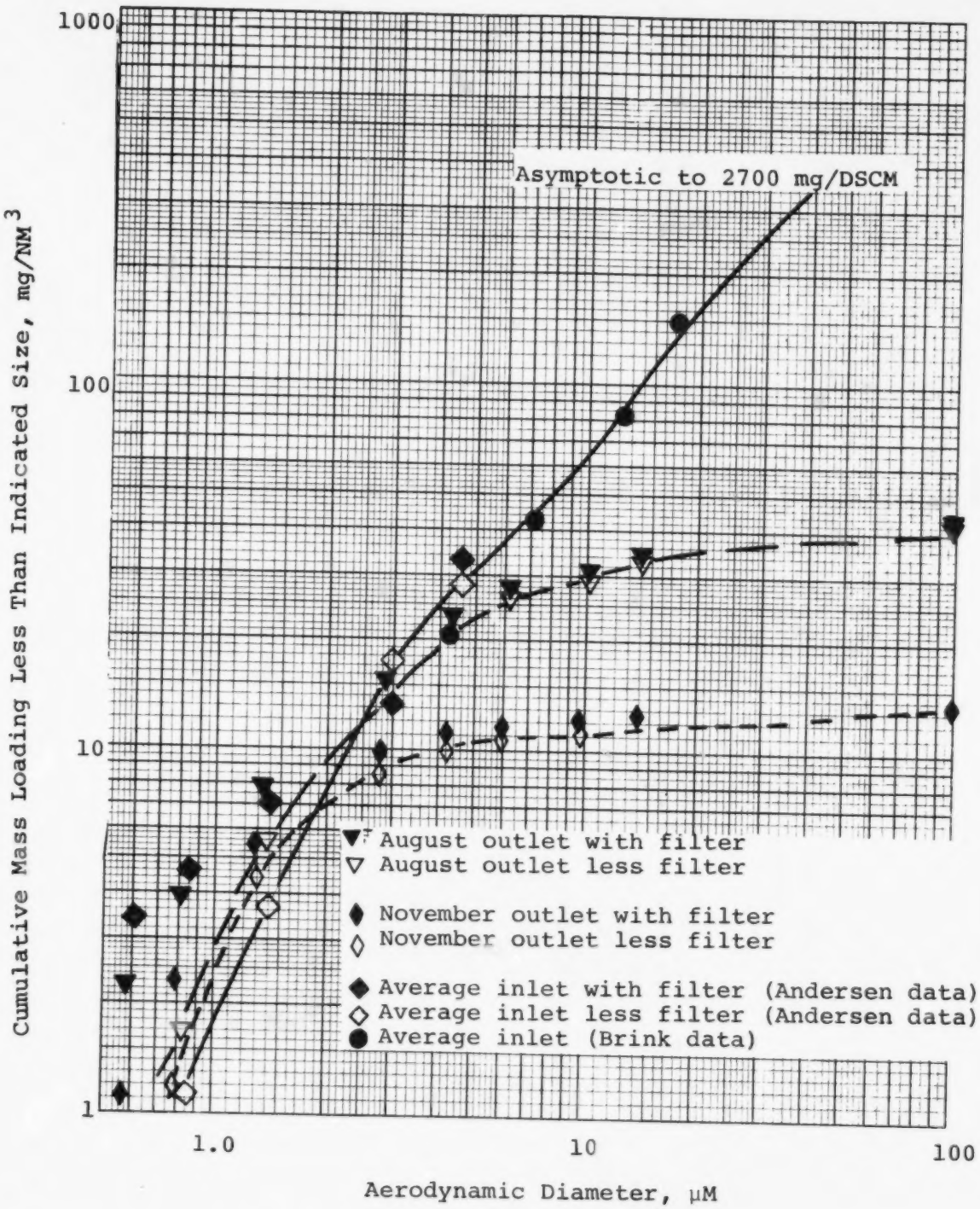


Figure 19. Inlet and outlet size distributions on a mass basis.

Table 7

Gravel Bed Filter Fractional Efficiencies As Measured With Andersen Impactors

Date	Inlet Gas Flow DSCM/min.	Inlet Temp. °C	Typical Bed Pres. Drop cm/H_2O	System Pres. Drop cm/H_2O	Backflush Period	Collection Efficiencies At Indicated Aerodynamic Diameters (μm)					
						.72	1.09	2.04	3.70	5.45	8.26
8/27	739	149	11.0	12.39	12 min.	-37	-68	32	60	91	97.7
8/28	917	181	13.6	17.3	12 min.	-146	-56	2	34	>71	
8/28	606	227	10.2	12.8	12 min.	4	-32	-42	19	>40	
8/29	1039	193	13.26	17.58	12 min.	13	-43	11	37	>67	
8/29	682	149	9.75	11.58	12 min.	-229	-56	25	67	>57	
11/4	732	166	6.4	9.6	6 min.	-104	-29	66	93	98.1	
11/5	856	152	9.1	14.0	6 min.	-31	0	78	94.1	96.3	

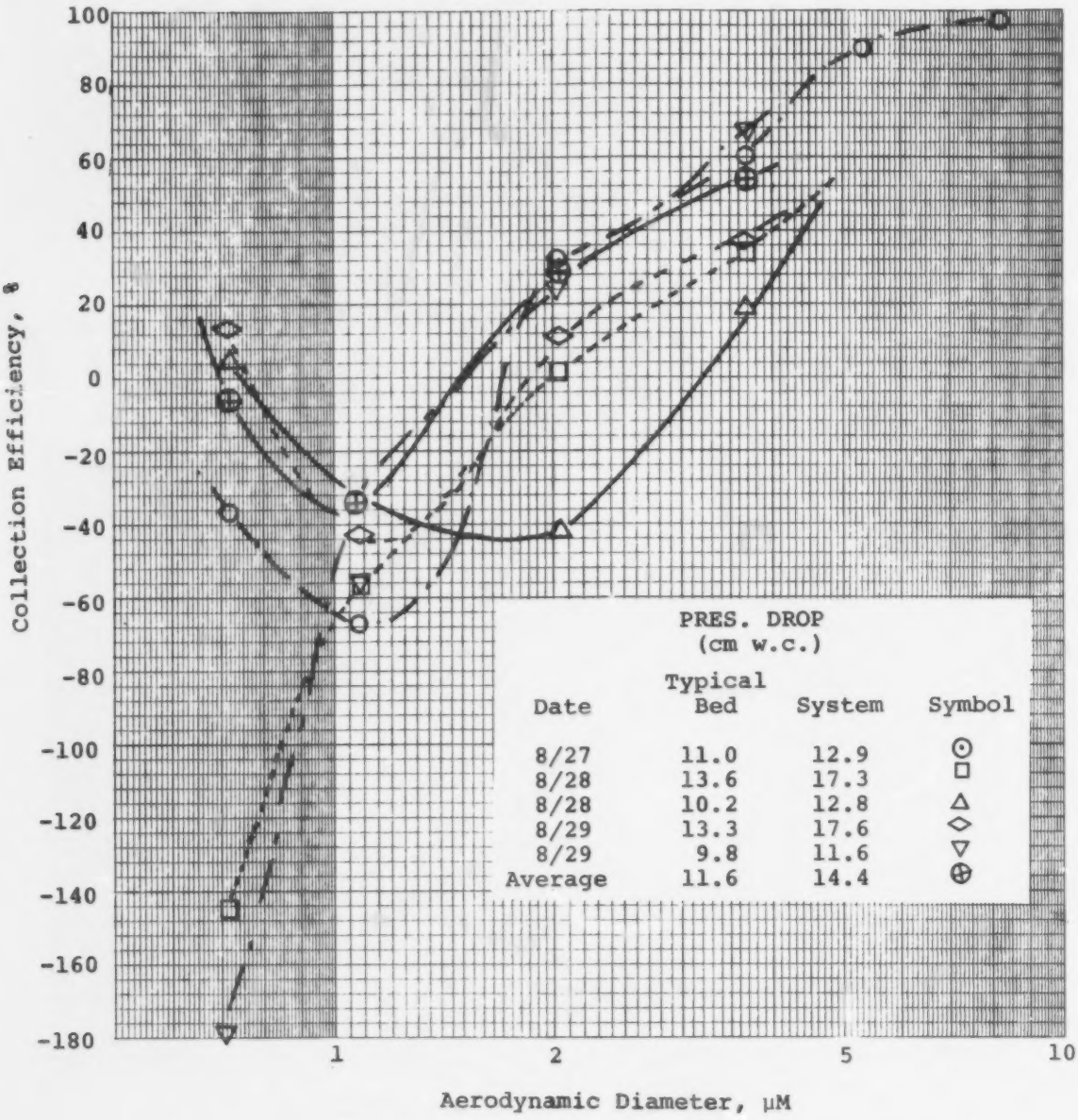


Figure 20. Fractional efficiencies as determined using Cascade impactors. August data.

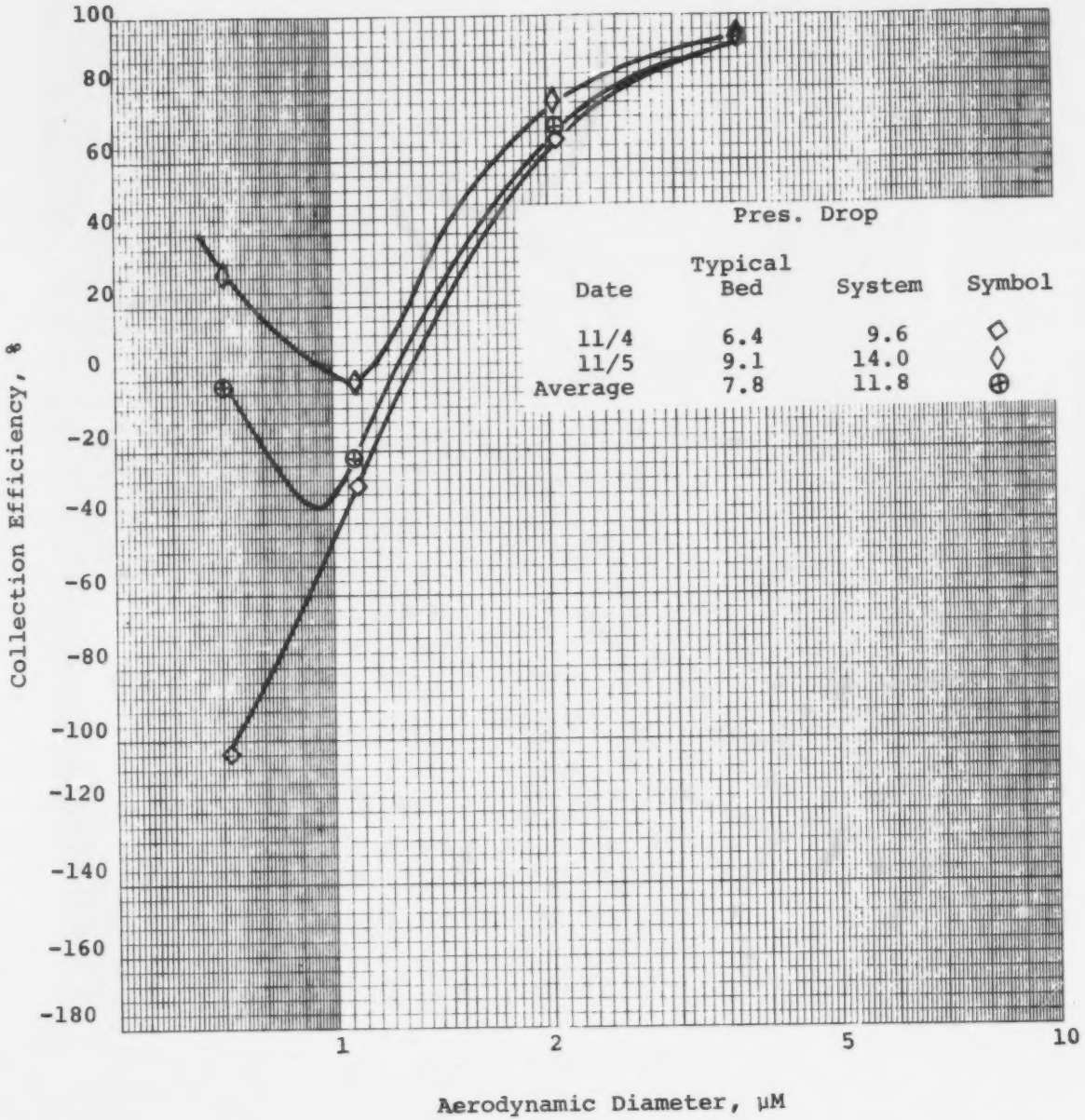


Figure 21. Fractional collection efficiencies as determined using Cascade impactors. November data.

using diffusional sizing techniques with General Electric Condensation nuclei counters for determining the various ultrafine particulate concentrations. Attempts were made to use a Thermosystems Model 3030 Electrical Aerosol Analyzer; however, rapid, random fluctuations in particulate concentrations and size distribution in this size range rendered the data from this method almost totally uninterpretable. In addition, a Gardner small particle detector (a manually operated CNC) was used throughout the tests as a crude monitor of the exit concentration and size distribution (using a variable supersaturation method) of ultrafine particles. Some inlet data were also obtained with the variable supersaturation method during the November tests. This instrument was used without dilution and much of the time the particulate concentration exceeded its range; however, it showed qualitative agreement with the data obtained using the more elaborate system in those instances during which they were at a common location. During the second (November) tests series, only the Gardner small particle detector was used to obtain data on ultrafine particulates. Accuracy in the measurements was limited by the rapid and frequent concentration variations and the efficiencies derived from these data are uncertain. However, the trends in the fractional efficiencies derived from the data are probably real and the fractions of the influent materials that penetrate the scrubber are believed to be generally correct to within a factor of two. The data obtained during the August test series are summarized in Table 8. The fractional efficiencies derived from the August data are shown in Figure 22. Also shown in Figure 22 are fractional efficiencies obtained from simultaneous measurements with the small particle optical counter and cascade impactors. The limited comparison data obtained during the November tests are given in Table 9. The efficiency for the +0.2 μm diameter particles in the November data appears to be substantially higher than was the case in August. It is unlikely that the collection efficiency in the 0.02 to 0.20 range was increased to a value as high as 80 to 90%; however, values of about 50% cannot be excluded.

Table 8

Ultrafine Particulate Data—August Test

	Cumulative Concentration Larger Than Indicated Size					
	Diffusional Method			Variable Supersaturation Method		
	Inlet	Inlet	Outlet	Outlet	Outlet	Outlet
Date:	8/26	8/27	8/28	8/26	8/27	8/28
Dia., μm						
.01	3.5×10^6	2.3×10^6	4.2×10^6	1.3×10^6	2.5×10^6	$>2.8 \times 10^6$
.02	2.0×10^6	1.8×10^6	1.7×10^6	7.9×10^5	2×10^6	$>2.8 \times 10^6$
.063	5.5×10^5	4×10^5	9.6×10^5			
.10	2.9×10^5	2×10^5	4.4×10^5			
.20	7×10^4	3×10^4	1.4×10^5	6.9×10^4	2.1×10^5	2.2×10^5

Table 9

Ultrafine Particulate Data—November Test

	Cumulative Concentration Larger Than Indicated Size*			
	Inlet		Outlet	
Date:	11/4	11/5	11/4	11/5
Dia., μm				
.01	$>2.8 \times 10^6$	$>2.8 \times 10^6$	$>2.8 \times 10^6$	$>2.8 \times 10^6$
.02	$>2.8 \times 10^6$	$>2.8 \times 10^6$	$>2.8 \times 10^6$	$>2.8 \times 10^6$
.20	2.1×10^5	3.0×10^5	1.8×10^5	3.5×10^5

*All by variable supersaturation method.

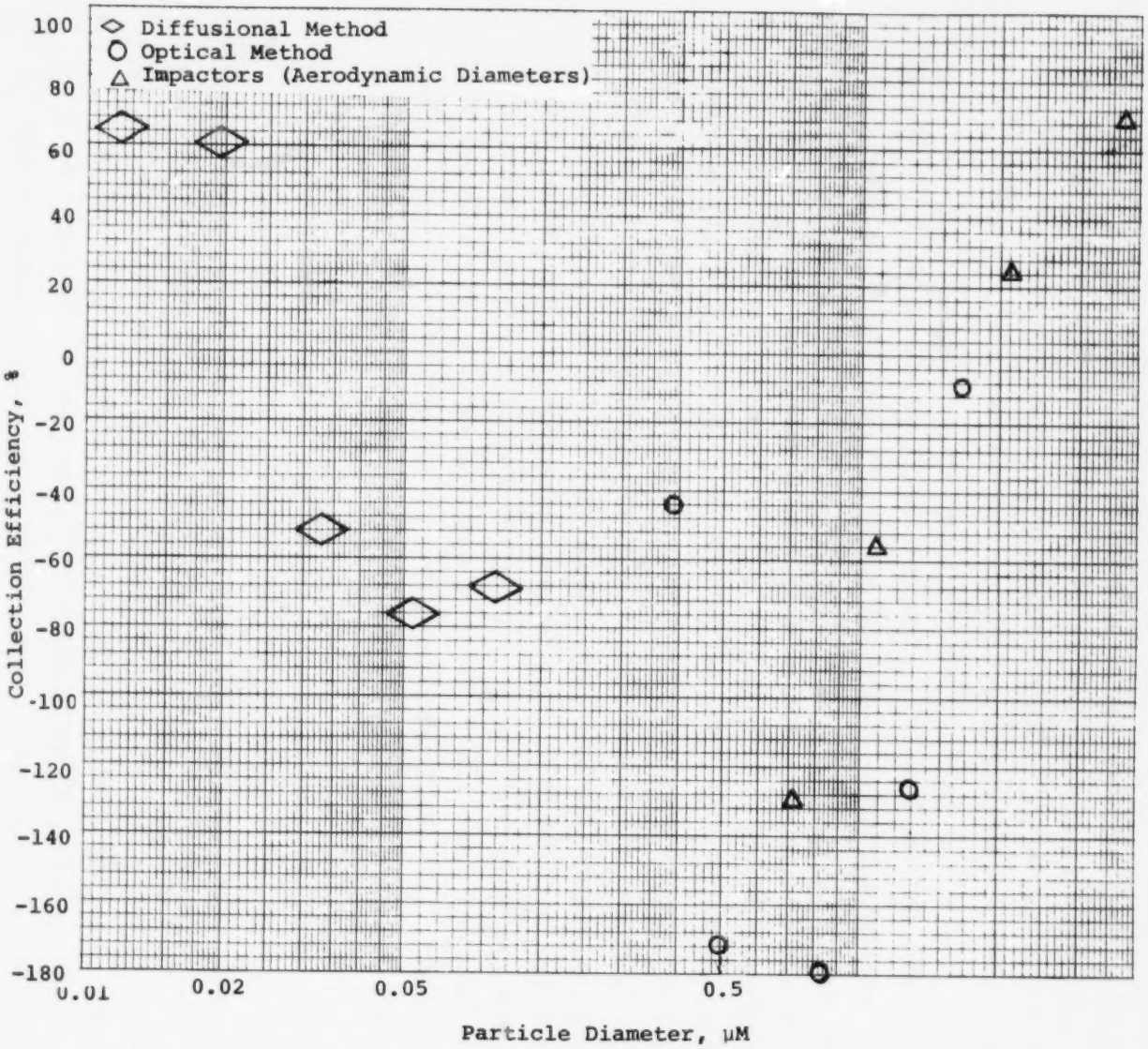


Figure 22. Fractional efficiencies as determined using diffusional, optical and inertial (Cascade impactor) methods. August data.

Two optical particle counters (Royco Model 225) were used during these tests. The first was used in conjunction with the ultrafine measurement system to obtain data on fractional efficiencies over the size range from 0.35 μm to 2.0 μm. The second was used only at the outlet of the gravel bed for obtaining real time information on relative changes in concentration of particles over the size range from 0.6 to 50 μm.

Inlet and outlet data with the small particle unit (0.35 to 2 μm) were obtained at the same times as were the ultrafine data. These data are thus not simultaneous sets and are subject to some of the same uncertainties as were the diffusional data. The inlet concentrations of particles in this size range were much more stable than the ultrafine concentrations, although still subject to large swings, and were probably dominated by entrained particulate from the clinker cooler and relatively unaffected by the ambient airborne particulate. The results of the measurements are given in Table 10. Fractional efficiencies determined from the optical data were shown in Figure 22, from which it can be seen that the negative fine particle collection efficiencies which were found from the impactor data were verified by the particle counter data.

The outlet data from both the small particle (0.3 to 2 μm) and large particle (0.6 to 50 μm) systems showed pronounced effects from the backflush cycle on emission rates. Typical examples of the time variations in the outlet particulate concentrations are shown in Figure 23, which represents an approximate 30-minute data segment obtained with the continuous monitoring large particle system during the August tests. A large "puff" followed by a slow decline in concentration occurs each time a clean module is put back on line (once every 12 minutes in this case). During the August test series, this 12-minute interval was varied for two special one-hour duration tests (one with 6-minute intervals, the other at 20-minute intervals) in order to allow some estimate

Table 10

Summary of Small Particle Optical Data

Concentration In Size Range (number/cc)

	Inlet		Outlet	
Date:	8/26	8/27	8/28	8/29
Averaging Time	(2.5 hrs)	(6 hrs)	(0.75 hrs)	(5 hrs)
Dia. Range, μm*				
.35-.43	5.0×10^2	7.46×10^2	14.3×10^2	9.0×10^2
.43-.58	3.4×10^2	4.7×10^2	14.0×10^2	13.0×10^2
.58-1.2	2.4×10^2	3.1×10^2	9.6×10^2	7.9×10^2
1.2-1.4	1.4×10^2	1.8×10^2	5.2×10^2	3.6×10^2
1.4-2.0	3.1×10^2	4.1×10^2	6.0×10^2	3.9×10^2

*Polystyrene equivalent diameter

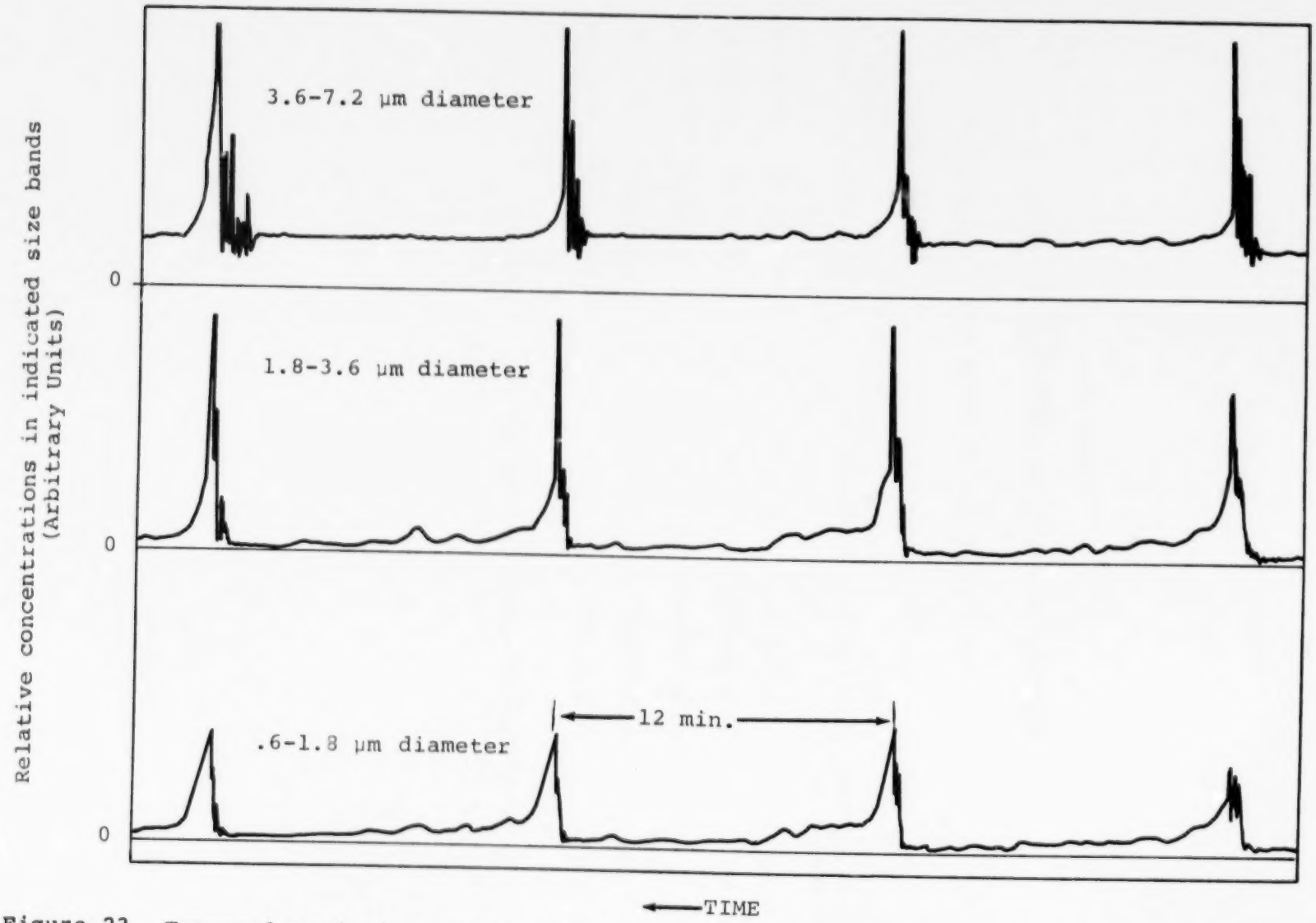


Figure 23. Temporal variations of particulate concentrations in three size bands at the outlet of the gravel bed filter system. Concentration units are arbitrary and not to the same scale for the three size bands.

to be made of the sensitivity of the gravel bed system to changes in operating conditions. The results of these tests are given in Table 11. During these one hour tests the system did not have time to come to a steady state condition, thus the results obtained during them may not well represent what might be obtained under steady-state conditions.

The November tests, which were made with the gravel bed system operating in a steady state condition with a 6-minute backflush time, showed a substantial improvement in performance as evidenced by the impactor data, in contrast to the rather small performance change observed during the one hour of operation with a 6-minute backflush cycle during the August tests.

The combined particle counter and cascade impactor data suggest that a large portion of the outlet fine particle emissions result from the breaking up of agglomerated fine particles during the raking operations in the cleaning of the off-line beds. The nature of the clinker cooler process is such that a substantial portion of the particulates lofted from the clinker bed by the cooling air could be agglomerates. A limited number of tests were run in November in order to explore the hypothesis that the negative efficiencies resulted from the breakup of agglomerates. These tests were made using only the realtime large particle system for monitoring the particulate concentrations in the gravel bed effluent gas stream. During these tests, clean beds were put on line without simultaneously backflushing another bed and dirty beds were backflushed with and without raking, with no clean bed being put on line.

The effects of these operations are illustrated in Figure 24 which shows strip chart recordings of the particle counter output during these tests. Events marked as "N" are the puffs normally observed when a new bed is placed in the backflush mode at which time a clean bed is simultaneously placed on line. Events marked "C" are the result of placing a clean bed on line without

Table 11

Effect of Backflush Interval On Particulate Emission Rate

Clean Module Insertion Interval (Minutes)	Relative Particulate Emission Rate In Size Bands Indicated (Arbitrary Units, 1 Hr. Avg.)		
	.6-1.8 μm dia.	1.8-3.6 μm dia.	3.6-7.2 μm dia.
6	36	42.5	37
12	31	39.2	40
20	25	38.5	52

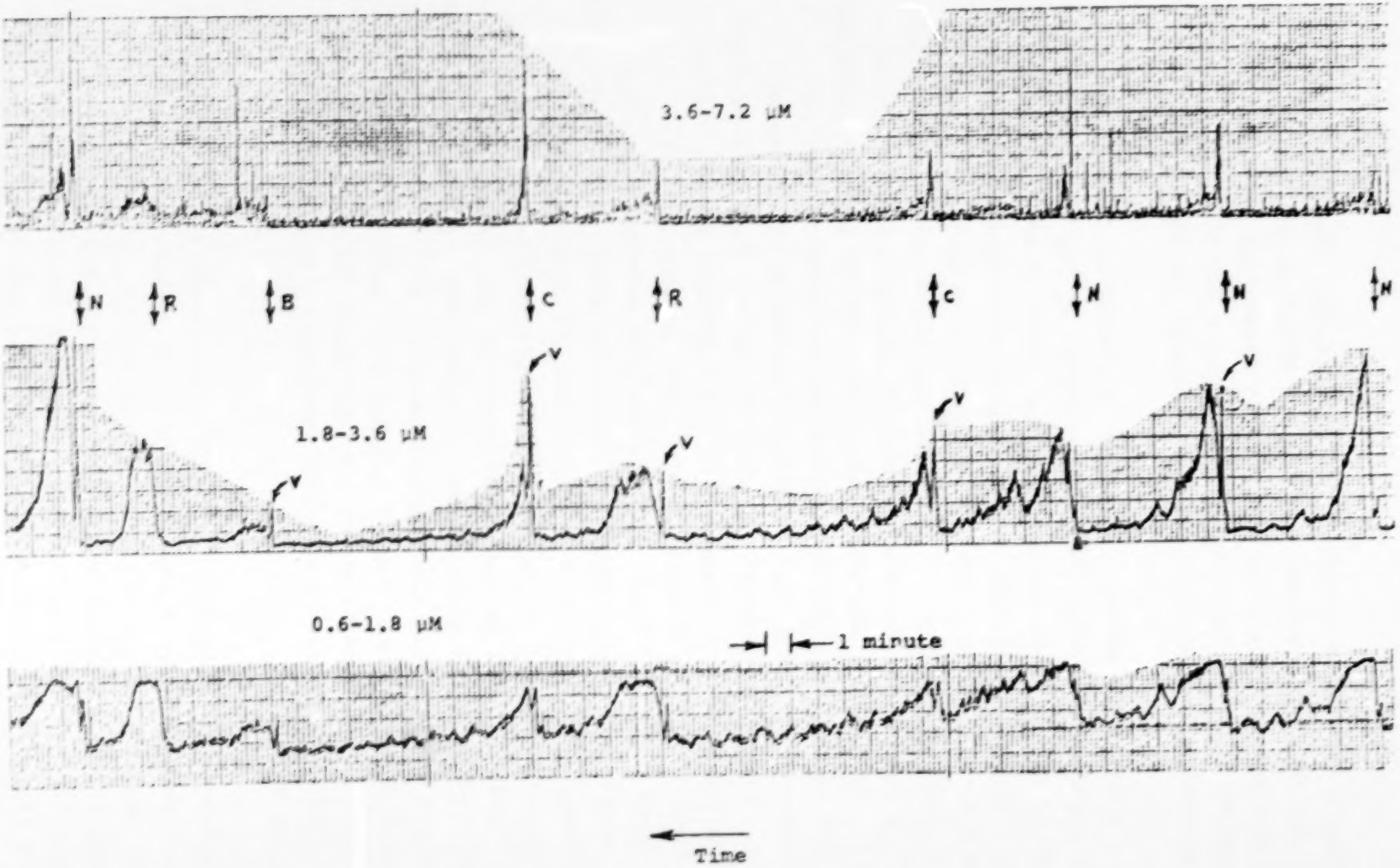


Figure 24. Relative emission rates in three particle size intervals during explorations of the effects of backwash operations on emissions. For explanation of symbols see text.

initiating a backflush cycle on another module. Events marked "R" are the result of raking a dirty bed which is being backflushed without simultaneously placing a clean bed on line. The event marked "B" resulted from backflushing a bed without raking it and without simultaneously placing a clean bed on line.

It appears that when a clean bed is placed on line, some residual particulate which has been loosened by the backwash operation is carried out directly to the stack causing a short duration, low amplitude puff. The valving operations also appear to produce a very short duration, concentration spike designated "V" in Figure 24. The initial burst of backflush air to a bed that is entering the cleaning cycle appears to create a puff of lesser amplitude. The raking process during the backwash process produces moderate to large amplitude concentration increases which last slightly longer than the time period during which the rakes are activated. The puffs labeled "B" and "R" result from particulate that was entrained in the backwash air and passed in the normal, forward flow direction through the on line, active, modules. It was therefore concluded that the backwash and raking cycle does in fact result in a large amount of deagglomeration of the particulate which has been collected by a bed, and that in some aerodynamic size ranges, the emissions resulting from the portion of this reentrained material that is not recollected by the active beds exceeds the rate at which particulate in those size ranges is carried into the system from the clinker cooler. Estimates of the percentages of total emissions in several size intervals resulting from the puffs were made from the data obtained with realtime monitors. These estimates are presented in Table 12 for the August and November tests.

Estimates of the fundamental collection efficiencies of the gravel beds were made by subtraction of relative contributions due to the puffs from the cascade impactor outlet data and recalculating fractional efficiencies based on the revised data. The results of these calculations are given in Table 13, together with the efficiencies calculated from the original data. These estimates indicate that the collection efficiency of the gravel

Table 12

Percentage of Emissions In Puffs

		Monitor #1		Monitor #2		
Date:	Size Interval:	.35-.6	.6-1.2	.6-1.8	1.8-3.6	3.6-7.2
August	(12 min Backwash)	29	51	55	70	62
November	(6 min Backwash)	N.A.	N.A.	(29)	67	(41)

N.A.: Not Available - Monitor #1 not used during November tests.
Values in parentheses are uncertain.

Table 13

Fractional Efficiencies From August and November Data With and Without Contribution From Cleaning Cycle

Date	Collection Efficiencies At Indicated Aerodynamic Diameters							
	.72 μm		1.09		2.04		3.7	
	with	w/o	with	w/o	with	w/o	with	w/o
8/27	-37	42	-68	13	32.1	69	60	72
8/28	-146	-3	-56	19	2.2	56	33.8	54
8/28	4.4	71	-32	31	-42	36	19	43
8/29	13	63	-43	26	11	60	37.2	56
8/29	-229	-38	-56	19	24.5	66	67	77
11/4	-104	(-53)	-29	(33)	66.2	90	93	97
11/5	-116	(-62)	15.5	(55)	73.8	92	94.8	98

Value in parentheses uncertain.

bed under the conditions of these tests was low for particles smaller than about 2 μm even without the effect of the puffs.

The difference between the August results and the November results is attributed primarily to the more frequent backwashing, which reduced the pluggage of the beds and resulted in lower operating pressure drops for the same gas flows. At the same time, as a result of the better cleaning of the beds, the residual dust concentration in the beds is lowered which in turn reduces the loading surge to the active beds when a bed is backwashed. The lower bed loadings also reduce the available dust which can be carried directly out to the stack when a bed is put back on line. It is also possible that with the beds in the highly plugged condition that existed during the August tests they may have become partially fluidized during the backwash cycle. This would enhance the deagglomeration process and perhaps result in the production of particulate from coarser particles by mechanical actions.

SECTION VII
VARIABLE THROAT VENTURI SCRUBBER

MANUFACTURER'S DESCRIPTION:*

The flue gas cleaning system (Figure 1) now in operation on the two Colstrip 360 MW units is unique in that a wet scrubbing system is used for both particulate and SO_2 control and captured ash provides the alkalinity for the SO_2 removal.

The system currently installed on the two 360 MW units 1 and 2 is illustrated in Figures 25 and 26. The hot flue gas leaving the boiler is cooled in the heat recovery air heater and enters the flue gas scrubbing system at about 300°F. Each scrubber module, as shown in simplified drawing in Figure 26, consists of a downflow venturi scrubber centered within an upflow spray tower contactor. The venturi is equipped with a variable throat to maintain constant pressure drop at variable loads. In the venturi the scrubbing liquid is finely dispersed by the high velocity flue gas and serves to efficiently wet and trap the particulate fly ash. In the spray tower the gas contacts a recycle spray of absorption slurry. The slurry from the venturi and the spray contactor is collected and held in the base of the scrubber and recirculated at an L/G rate of 15 for venturi and 18 for the absorber spray. An agitator in the scrubber base serves to maintain suspension of the fly ash and solid reaction products. Slurry is bled from the recycle to maintain a 12% suspended solids concentration. Slaked quick lime is added as lime slurry

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*Taken from a paper by C. Grimm, J. Z. Abrams, W. W. Leffmann, I. A. Raben, and C. Lamatia. Presented at the 1977 National Meeting of the AIChE.

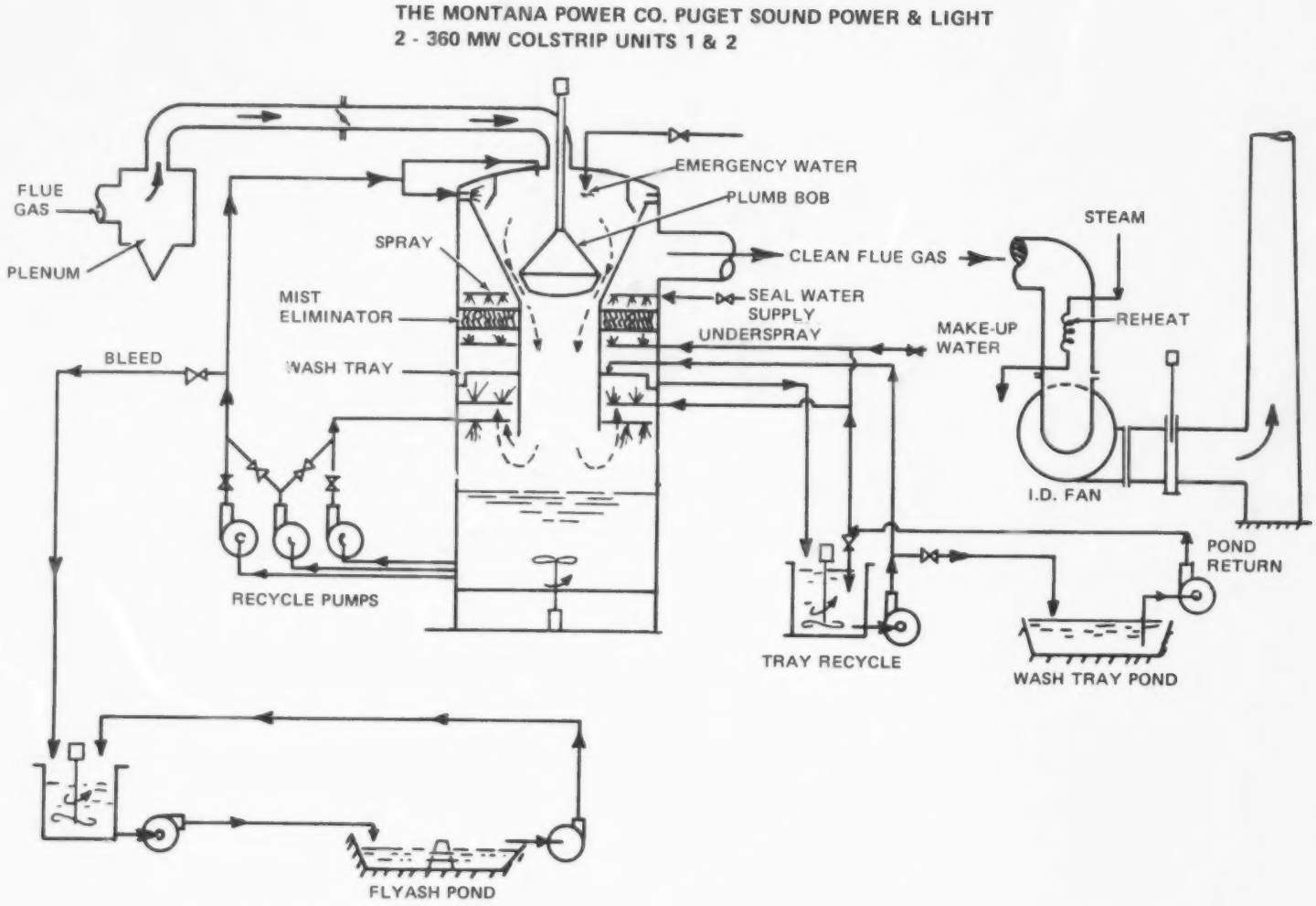

Figure 25. The Montana Power Co.-Puget Sound and Light Colstrip Units 1 and 2-(360 MW Each) Flue Gas Cleaning System.

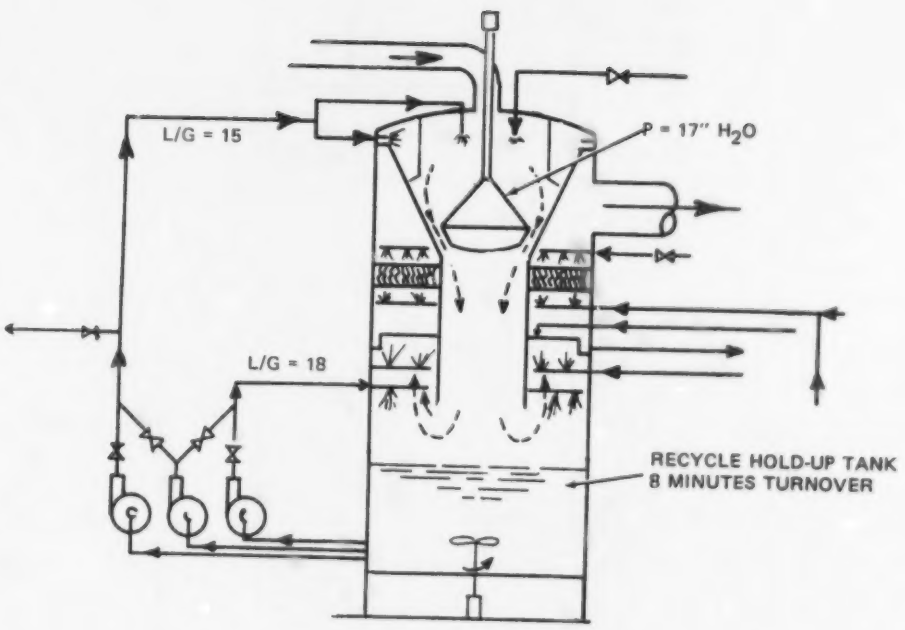

Figure 26. Colstrip Scrubber Module

only if needed to augment the fly ash alkali and maintain the desired slurry pH.

Each scrubber module is designed to clean 120 MW of equivalent gas flow under normal conditions and 144 MW under emergency conditions. (i.e., when one module is down, the two in operation will clean the amount of flue gas generated at 80% of boiler design load.)

The treated gas leaving the spray section passes through the water wash tray which serves to trap and dilute the entrainment. The gas leaving the washtray passes through a chevron demister followed by a mesh pad demister and leaves the absorption section water-saturated and cooled to the saturation temperature of about 120°F.

To preclude condensation in the fan and stack, and improve the gas buoyancy, the cooled gas from the scrubber is reheated to 50-75°F by a steam-heated exchanger. The warmed gas then passes through the dry induced draft fans and is discharged to the atmosphere from the top of a 500 foot stack.

As shown in Figure 25 the slurry discharged from the absorption loop is passed to an intermediate retention pond where the solids settle and from which the clarified water is returned to the absorption system. At intermittent intervals (currently only during the warm summer months), a floating dredge is used to reclaim the settled solids from the intermediate settling pond and transport them as a 30% slurry by pipeline to the remotely located permanent disposal pond. Decanted water (supernate) from the disposal pond is returned, also intermittently, through the same slurry pipeline to the intermediate pond for recycle to the absorption system. No stabilization of the sludge is required and a closed water loop is maintained.

Fresh water is added to the absorption system in an amount equivalent to that evaporated into the warm gas stream plus that

retained in the waste sludge. This fresh makeup water is introduced to the system as dilution water for minimizing the calcium saturation level in the mist eliminator washwater. This washwater is trapped by and withdrawn from the washtray and circulated to a small pond where entrained solids are separated. A portion of the water from this pond is returned and used to wash the undersurface of the washtray. Another portion of the flow is diluted with the fresh makeup water, and used for bottom wash of the mist eliminator.

The scrubber has been free of scale while the pH of the recycle liquid remains in the expected range. Corrosion problems in the reheater and demister plugging have not been experienced with the installation.

A detailed chemical analysis of the fly ash (see Table 14) revealed that it contained alkali metal oxides in an amount theoretically sufficient to react with and absorb the sulfur dioxide produced by the coal combustion. Laboratory experiments simulating absorption conditions revealed that this alkalinity was only usable under low pH absorption conditions (<5.6). It also revealed that absorption under these low pH conditions would result in extensive oxidation of the absorbed SO_2 producing calcium sulfate rather than calcium sulfite as the predominant reaction product.

Laboratory tests were conducted by Bechtel to determine the process conditions under which the alkalinity of the fly ash could be utilized while at the same time accommodating the scaling potential of the calcium sulfate. The conditions selected were a pH of 5 to 5.6, low enough for alkali utilization and high enough for adequate SO_2 absorption capability. The other, and perhaps the key operating factor, was the use of a high level of suspended solids in the absorption slurry (12 to 15% by weight, of which some 3-4% is calcium sulfate formed in the absorption). This provided a high concentration of calcium sulfate seed crystals

Table 14

Fuel And Ash As Described
In Specifications

COAL:	Average, As Received
Moisture	23.87%
Volatile Matter	28.59%
Fixed Carbon	38.96%
Ash	8.59% (Max. 12.58%, Min. 6.1%)
Heating Value	8843 Btu/lb. (Min. 8162 Btu/lb.)
Sulfur	.777% (Max. 1.0% Min. 0.4%)

ASH: (Estimated composition, sulfur trioxide-free basis)

SiO_2	41.60%
Al_2O_3	22.42%
TiO_2	0.79%
Fe_2O_3	5.44%
CaO	21.90%
MgO	4.95%
Na_2O	0.31%
K_2O	0.13%
P_2O_5	0.41%
(balance unidentified)	

Later fly ash data varies slightly from above as follows:

LEACHED IN H_2O (1% Fly Ash)

pH	11.8
Conductivity	4.150
Total Dissolved Solids	930 ppm
Calcium	396 ppm
Magnesium	0 ppm
Chloride	15 ppm
Sulfate (SO_4=)	30 ppm

LEACHED IN HCl

% Acid insolubles (SiO_2)	57.59
% Calcium as CaO	22.00
% Magnesium as MgO	1.27
% Aluminum As Al_2O_3	15.59
% Iron as Fe_2O_3	4.97
% Sulfate as SO_4	0.71
% Carbonate as CO_3	0.70

to promote desupersaturation. A long residence time for the recycle slurry in a stirred tank external to the scrubber was also proposed to ensure alkali utilization and to provide crystallization of calcium sulfate under controlled and non-scaling conditions. A slurry holdup of 8-10 hours was selected based on bleed rate.

The above two conditions, i.e., low slurry pH and long contact with the oxygen-containing flue gas, provided substantially complete oxidation. This high oxidation was shown to improve the disposal characteristics of the waste sludge produced.

TEST RESULTS:

The tests were conducted on one of the three identical scrubber modules which are operated in parallel to control SO_2 and particulate emissions from the power boiler. The three modules are independently controlled with respect to liquor flows and venturi pressure drop. Pressure drops across the venturis are regulated by adjusting the position of the "plumb bob" shown in Figure 27, thereby increasing or decreasing the cross sectional area of the venturi throat. Throughout these tests, with the exception of one brief period, the pressure drop across the venturi on the module being tested was held at 46 ± 2 cm w.c. Gas temperatures at the scrubber inlet ranged from 129°C to 137°C. The scrubber exit gas temperatures ranged from 57°C to 60°C and temperatures at the outlet test plane ranged from 94°C to 99°C. The temperature rise between the scrubber exit and the outlet mass sampling location results from a flue gas reheat system and the action of the fan, both of which are located between the scrubber outlet and the sampling plane. The gas flow handled by the scrubber throughout the tests was approximately 130 DNCM/sec (280,000 DSCFM). The venturi sections have a designed pressure drop range of 30.5 cm (12 in.) to 50.8 cm (20 in.) w.c. with a nominal operating pressure drop of 43 cm (17 in.) w.c. Design L/G rates are 2.0 ℓ/m^3 (15 gal/1000 CF) in the venturi section and 2.41 ℓ/m^3 (18 gal/1000 CF) in the absorber section.

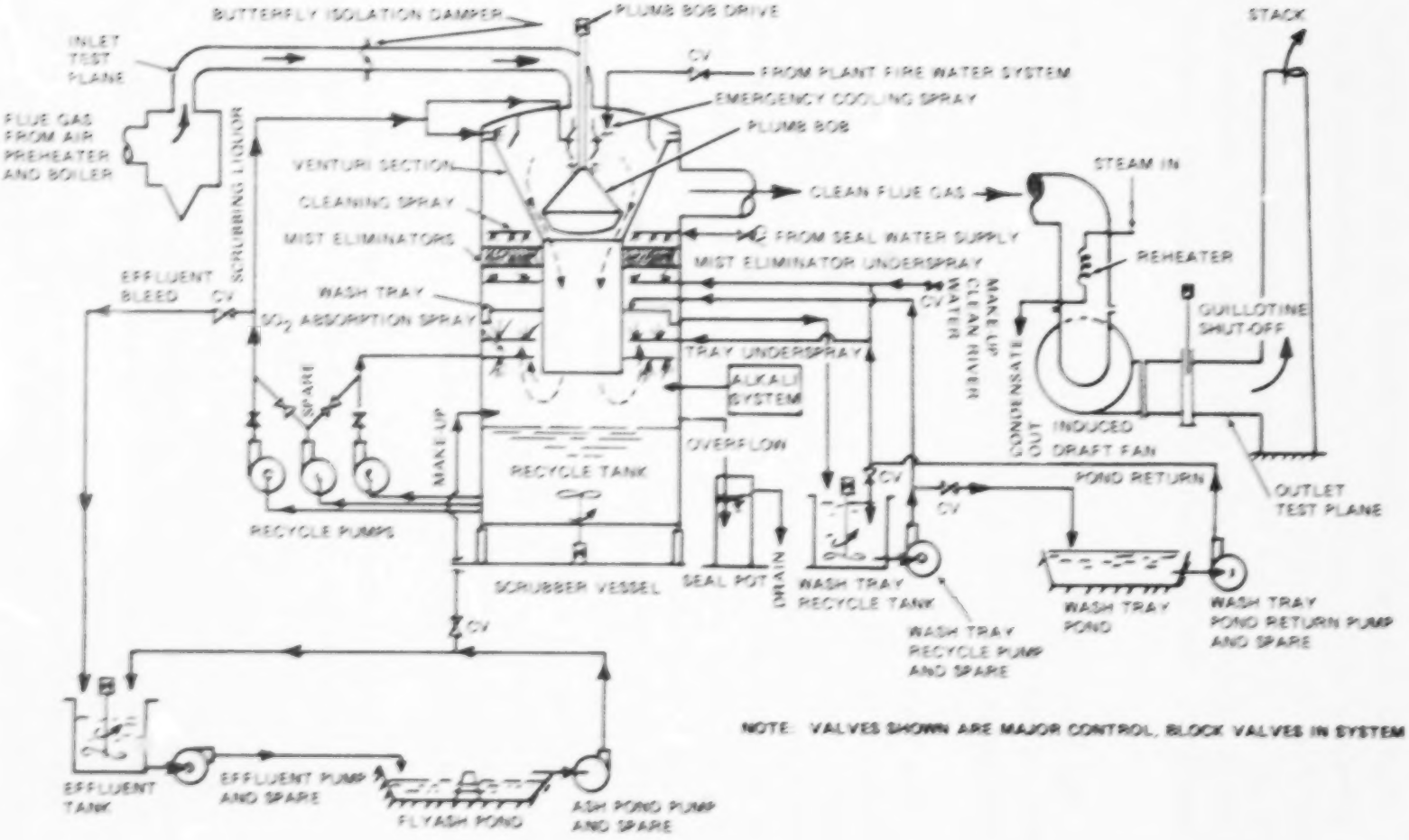

Figure 27. *Simplified scrubber flow diagram.*

High alkali metal oxides content in the fly ash permitted the scrubber to be designed to use a recirculating slurry of flyash for SO_2 removal. The system operates with a slurry having a pH of 5.0 to 5.6 containing 12 percent solids by weight. The pH is controlled by the addition of small amounts of lime as required. The scrubber design parameters are given in Table 15.

The data obtained by Method 17 testing are summarized in Tables 16 and 17. The overall collection efficiencies for each of the pairs of tests are given in Table 18.

The impactor data are summarized in Figures 28 through 31. Figures 28 and 29 present averaged inlet and outlet size distributions, respectively, on a cumulative percentage (by mass) basis versus aerodynamic particle diameter, while Figures 30 and 31 show the same data on a cumulative mass concentration basis. Figure 32 shows the fractional efficiency curve as a function of aerodynamic particle diameter as derived from the inlet and outlet data that were presented in the previous figures. The fractional efficiency curve is shown later in Figure 35 on a Stoke's diameter basis (ρ = 2.30) together with the efficiency curves derived from the ultafine particulate data. The scrubber was operating at a venturi pressure drop of about 48 cm w.c. throughout the impactor test periods.

Inlet size distributions obtained using optical and electrical mobility methods are shown in Figure 33. Outlet size distributions on a similar basis are shown in Figure 34 for the normal scrubber operating conditon (48 cm w.c. venturi pressure drop). Figure 35 shows the fractional efficiencies for ultrafine particles. Also shown in Figure 35 are the fractional efficiencies as a function of Stoke's diameter, obtained from the impactor data.

The scrubber was operated at venturi pressure drops of 31, 36, 41, 47, and 51 cm w.c. for a brief period at each condition

Table 15

Design Parameters For The CEA Variable Throat Venturi Scrubber
(Colstrip Application)

Venturi Pressure Drop	43.2 cm w.c. (17 in.)
Venturi L/G	2 ℓ/m (15 gal/1000 ACF, saturated)
Absorption Spray L/G	2.41 ℓ/m (18 gal/1000 ACF, saturated)
% suspended solids in recirculating slurry, by weight	12%
Residence time in the recycle tank	8 minutes
Gas velocity in mist eliminator zone	2.65 m/sec (8.7 ft/sec)
Wash tray pressure drop	9.65 cm w.c. (3.8 in.)
Mist eliminator pressure drop	2.5 cm w.c. (1 in.)
Reheat pressure drop	5.6 cm w.c. (2.2 in.)
Total system pressure drop (including reheat)	64.8 cm w.c. (25.5 in.)
Total scrubber pressure drop (less reheat)	55.4 cm w.c. (21.8 in.)

Table 16

CEA Variable Throat Venturi Test
Inlet Mass Data

Run Number	1	2	3	4	5	6
Date	5-16-77	5-17-77	5-18-77	5-18-77	5-19-77	5-19-77
Time	1715	1455	1235	1545	0825	1245
Moisture, %	10.30	11.62	10.25	10.87	11.86	12.26
Gas Temperature, °C	134	132	129	129	137	133
°F	274	269	265	264	278	272
Volumetric Flow, m^3/sec	208.6	201.3	233.9	236.9	238.8	227.6
ACFM	442,000	426,500	495,500	502,000	506,000	482,300
Volumetric Flow, DNCM/s	116.5	110.3	131.9	132.9	130.1	124.4
DSCFM	247,700	233,600	279,500	281,500	275,600	263,500
Concentration, grams/ACM	2.0184	2.8325	3.3097	3.4820	3.5145	3.5829
Concentration, grams/DNCM	3.6145	5.1701	5.8663	6.2079	6.4512	6.5546
Isokinetic, %	107.62 *	105.85 *	104.23	108.62	103.79	103.56 **

* Used points for 10 ft. stack
**Test cut short by 3 points (6 minutes) due to boiler shutdown.

TABLE 17

CEA Variable Throat Venturi
Outlet Mass Train Data

Run Number	1	2	3	4	5	6
Data	5-16-77	5-17-77	5-18-77	5-18-77	5 19-77	5-19-77
Time	1700	1315		1500	830	1300
Moisture, %	14.01	19.45	17.37	16.53	18.70	18.15
Gas Temp.						
°C	99.4	94.4	96.1	96.1	96.1	96.1
°F	211	202	205	205	205	205
Volumetric						
Flow, M^3/s	194.9	224.8	238.9	273.8	237.8	239.3
ACFM	413,000	476,200	506,200	579,300	503,800	507,000
Volumetric						
Flow, DNM^3/s	118.4	128.4	140.0	162.1	137.3	139.3
DSCFM	250,800	171,100	196,700	343,500	290,900	295,200
Concentration, mg/ACM	26.09	25.86	19.66	21.52	24.23	19.23
Concentration, mg/DNCM	42.79	45.31	33.58	36.28	41.90	33.04
Isokinetic, %	105.71	113.99	106.40	106.76	103.91	104.66

TABLE 18

CEA Variable Throat Venturi Scrubber Efficiences
From Mass Train Data

Date	Efficiency (%)
5-18-77	99.43
5-18-77	99.42
5-19-77	99.35
5-19-77	99.50

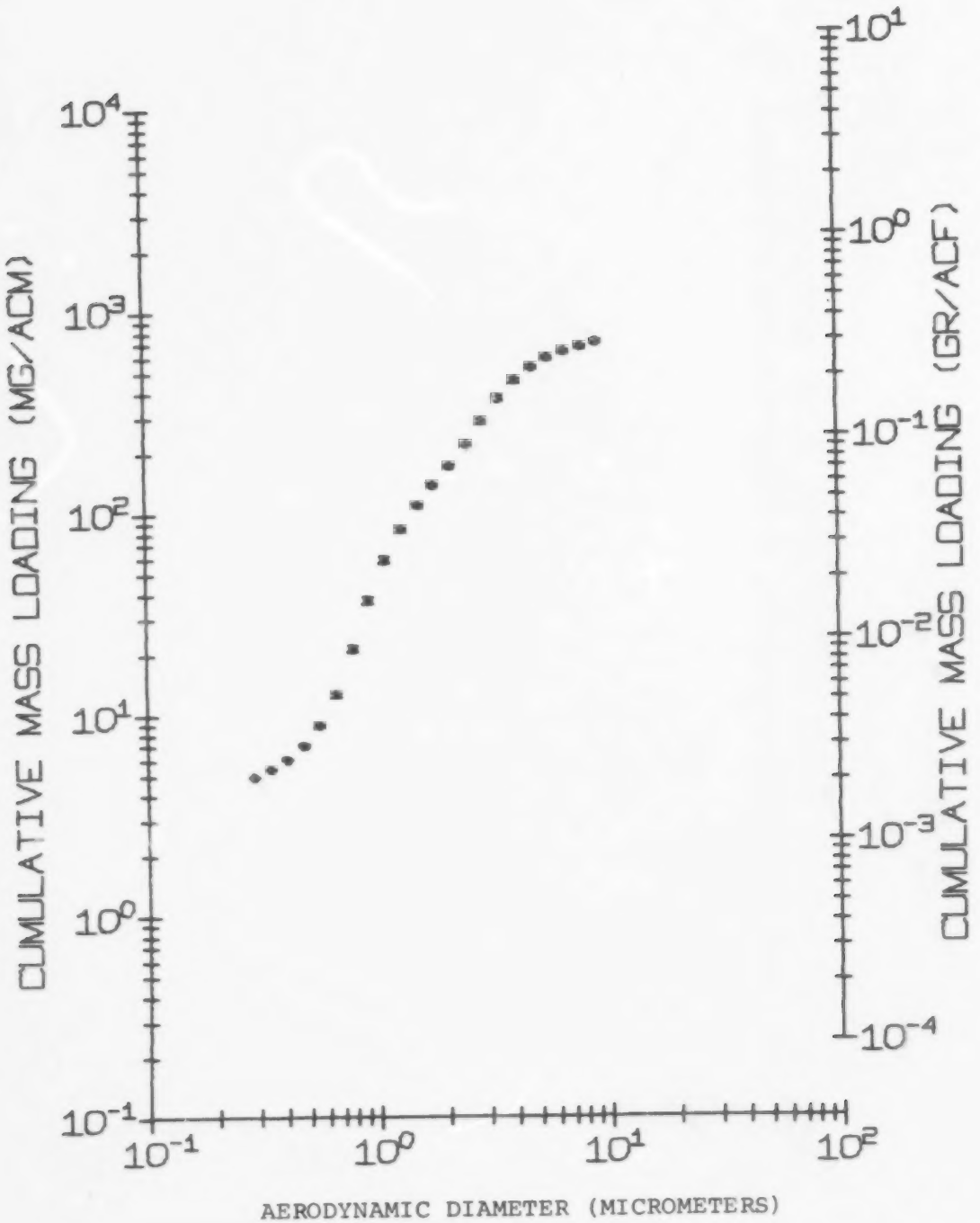

Figure 28. Average inlet particle size distribution on a cumulative mass concentration basis from cascade impactor data.

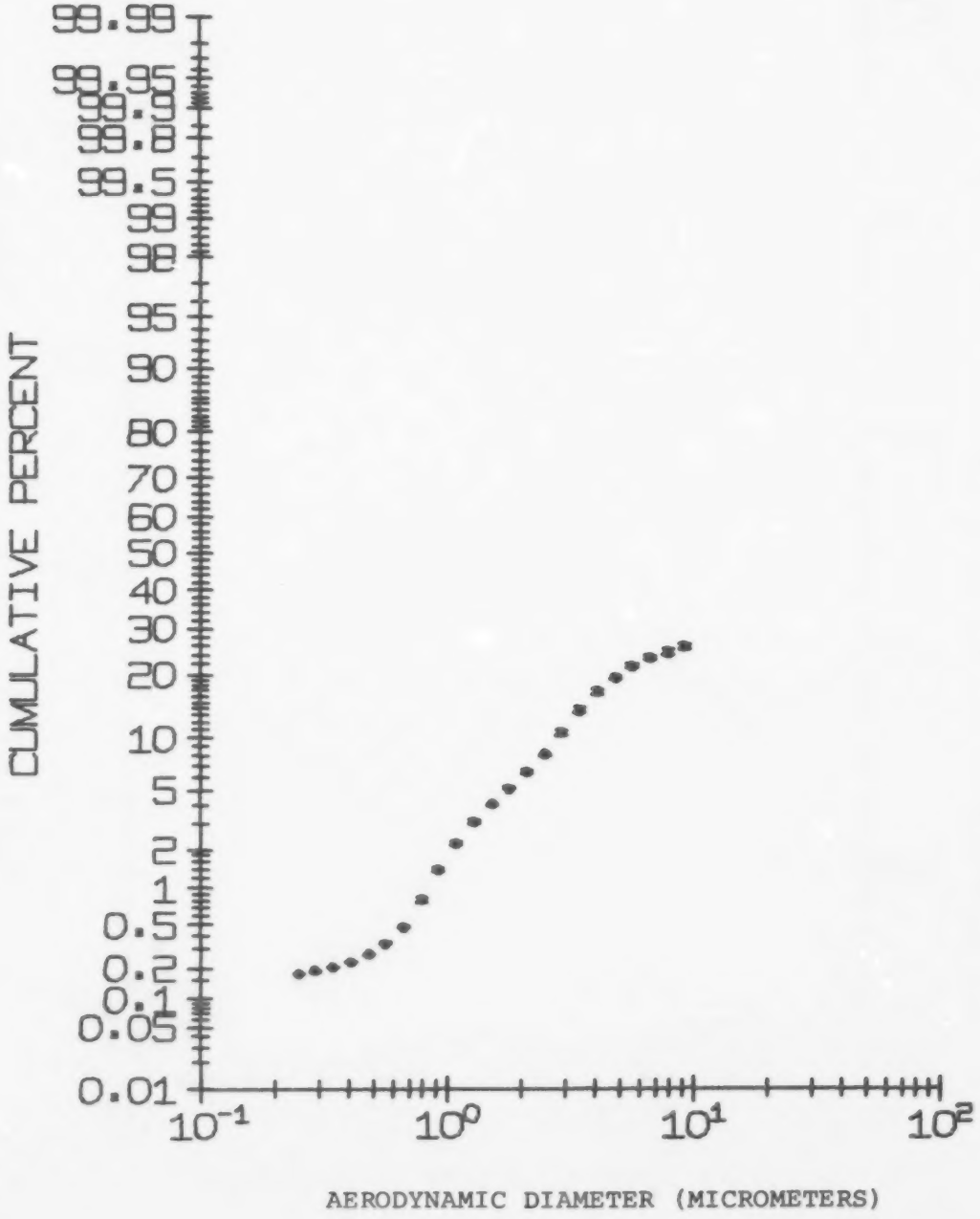

Figure 29. Average inlet particle size distribution from cascade impactor data on a cumulative percent by mass basis.

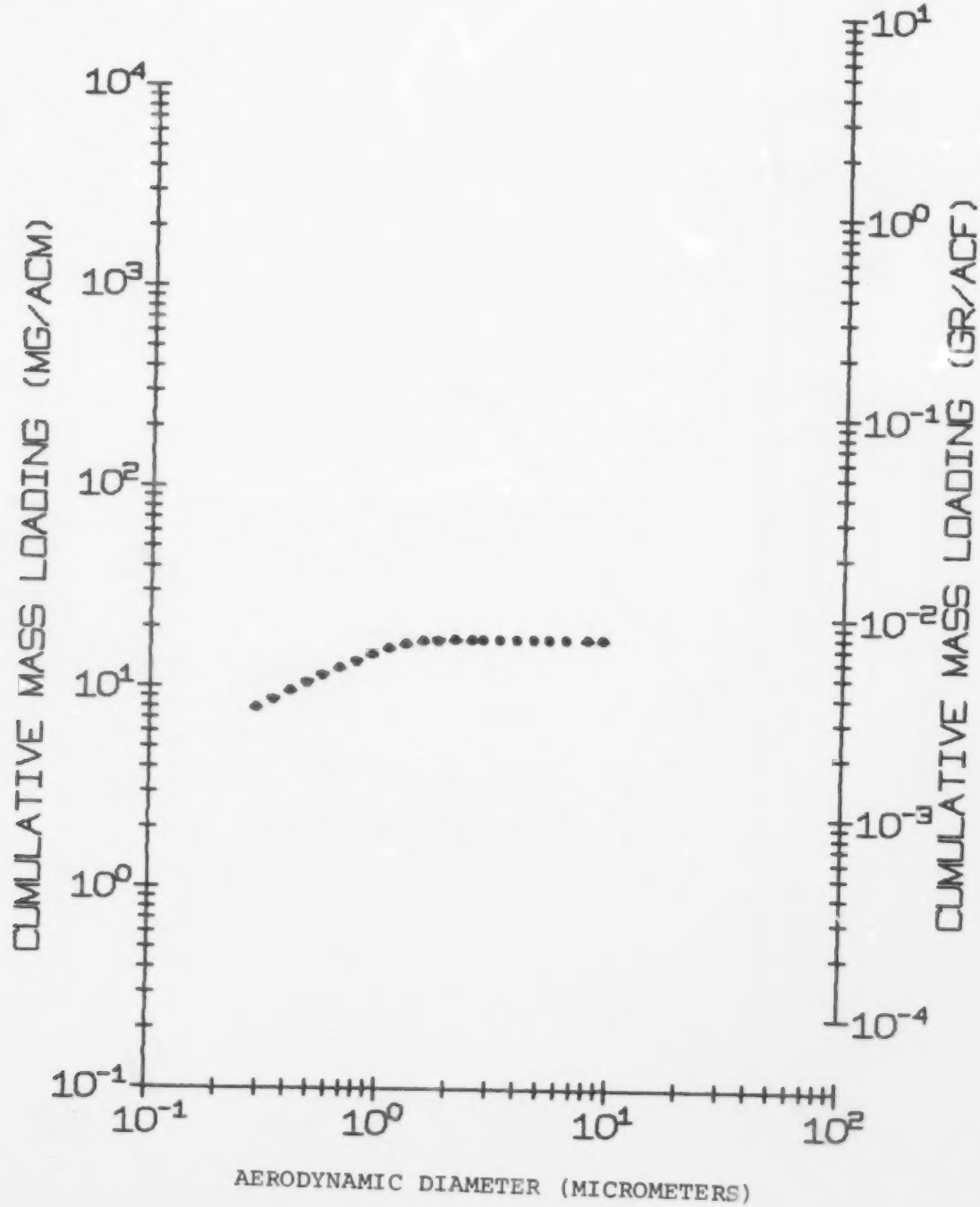

Figure 30. Average outlet particle size distribution on a cumulative mass concentration basis from cascade impactor data.

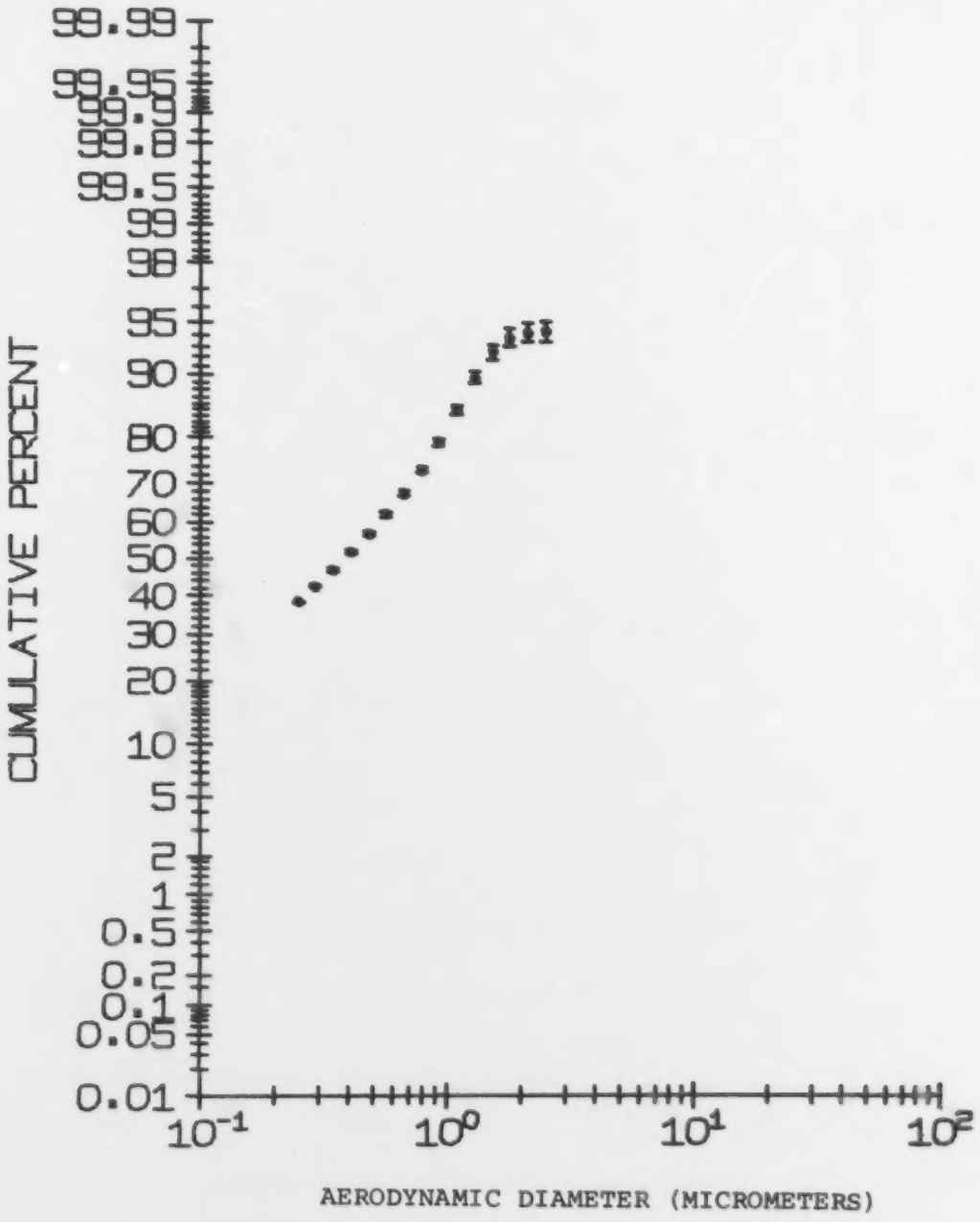

Figure 31. Average outlet particle size distribution from cascade impactor data on a cumulative percent by mass.

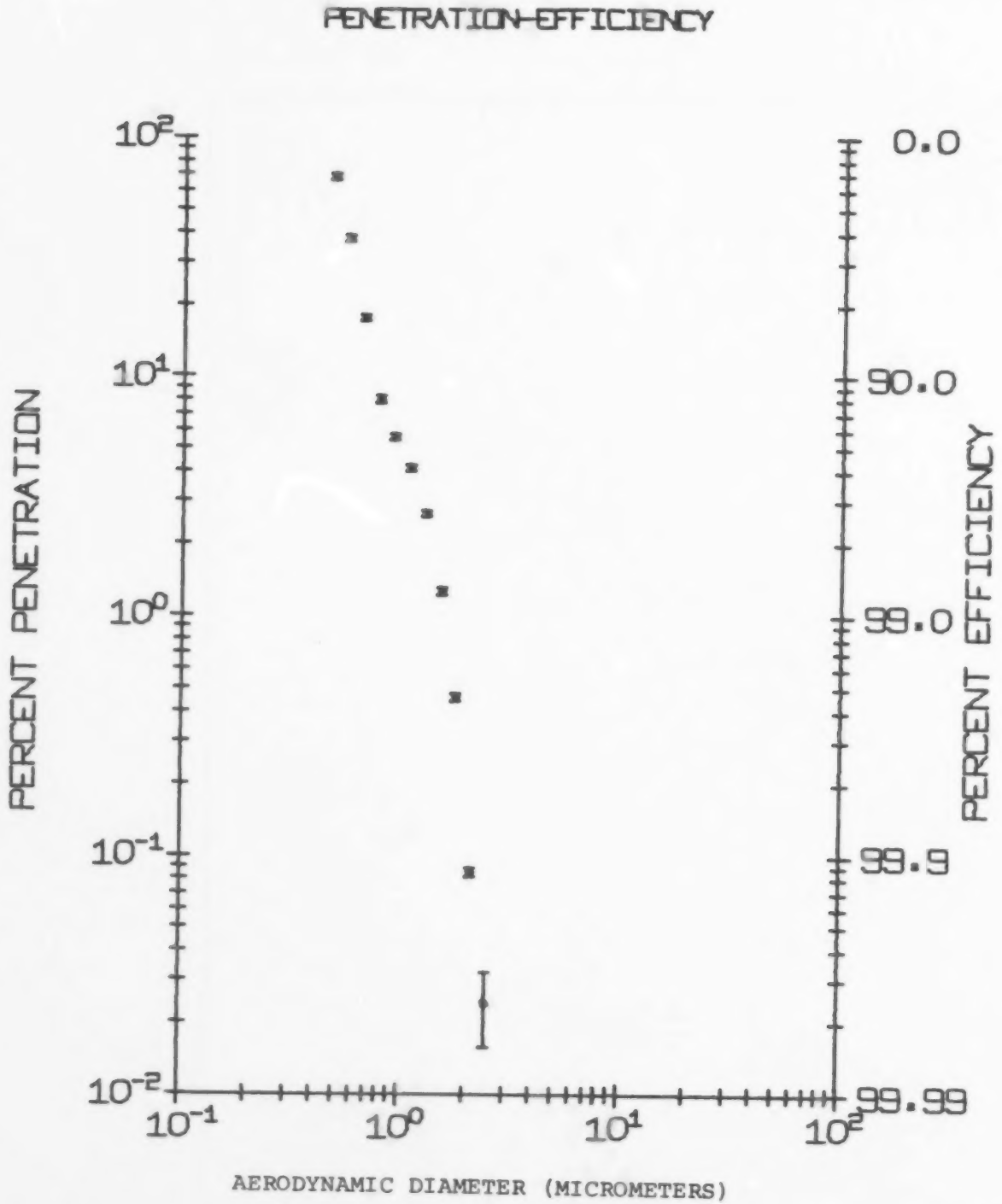

Figure 32. Fractional efficiency curve on an aerodynamic particle diameter basis for the CEA variable throat venturi scrubber operating at a venturi pressure drop of 48 cm (19 in.) w.c..

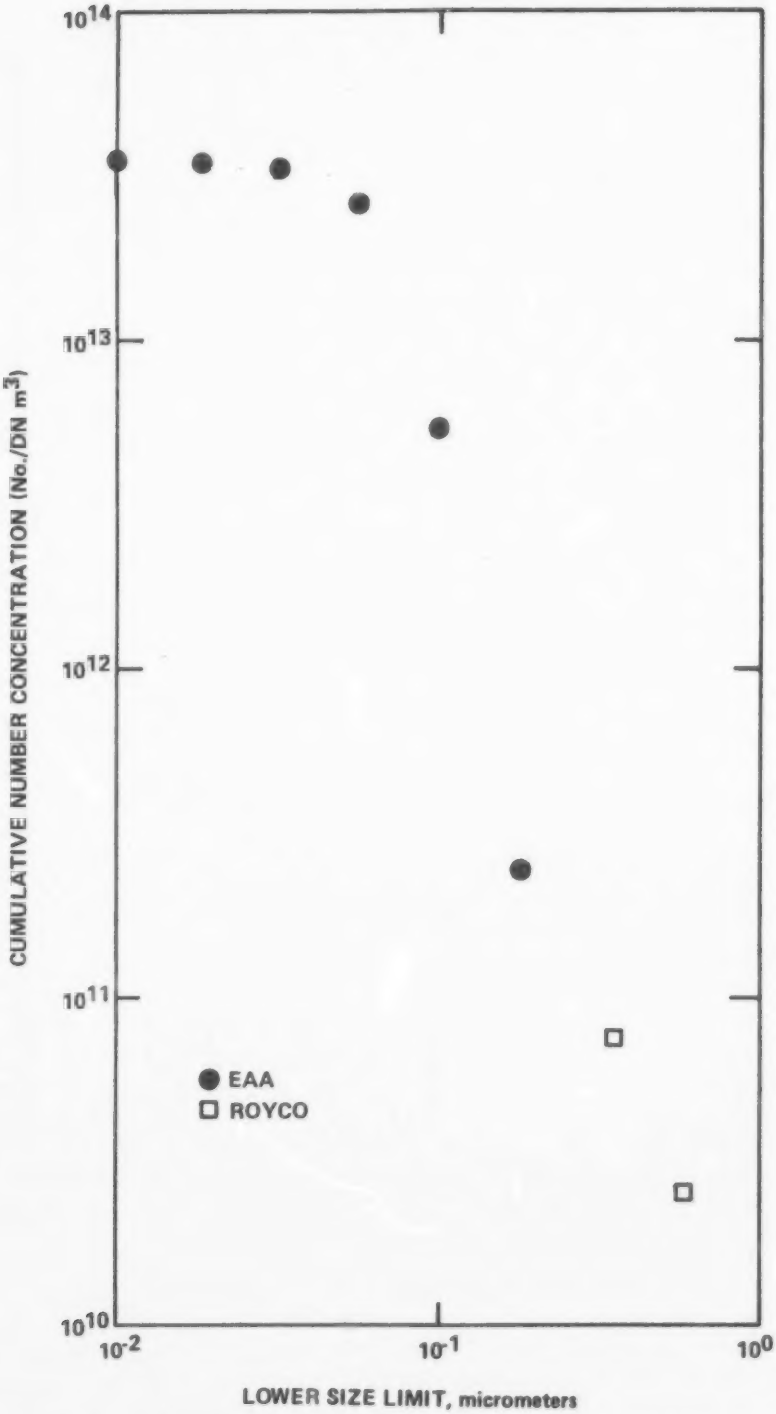

Figure 33. Scrubber inlet particle size distribution from electrical aerosol analyser data.

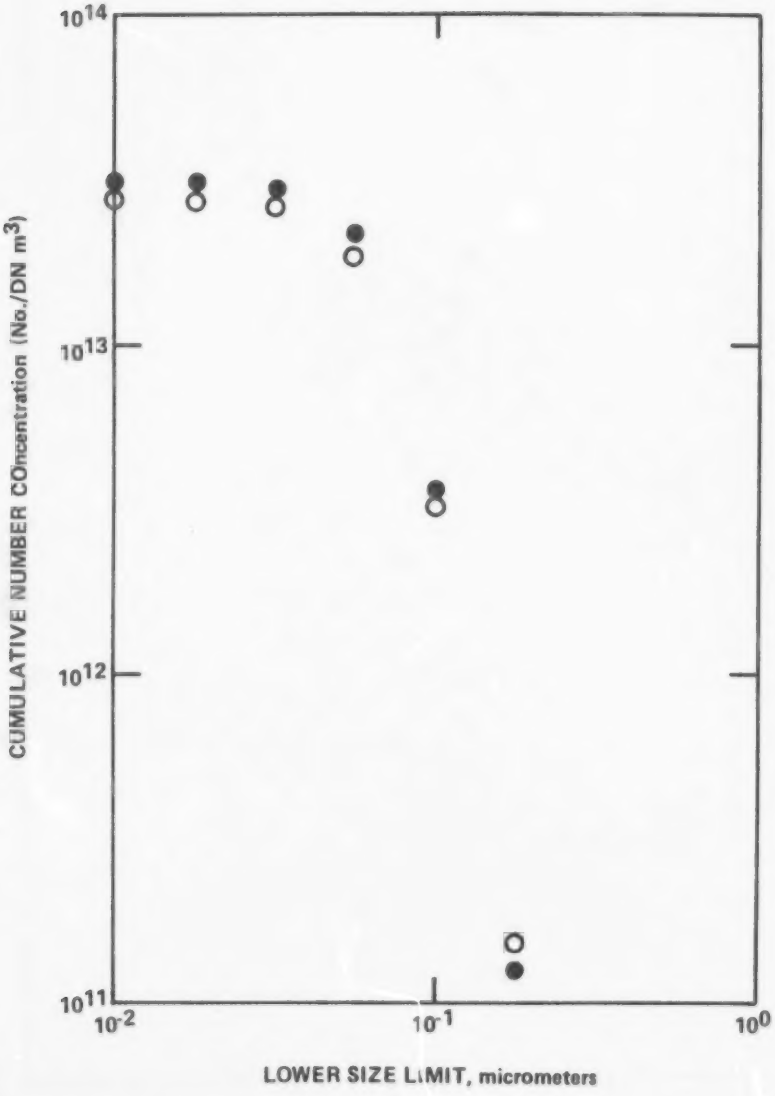

Figure 34. Scrubber outlet particle size distribution from electrical aerosol analyser data.

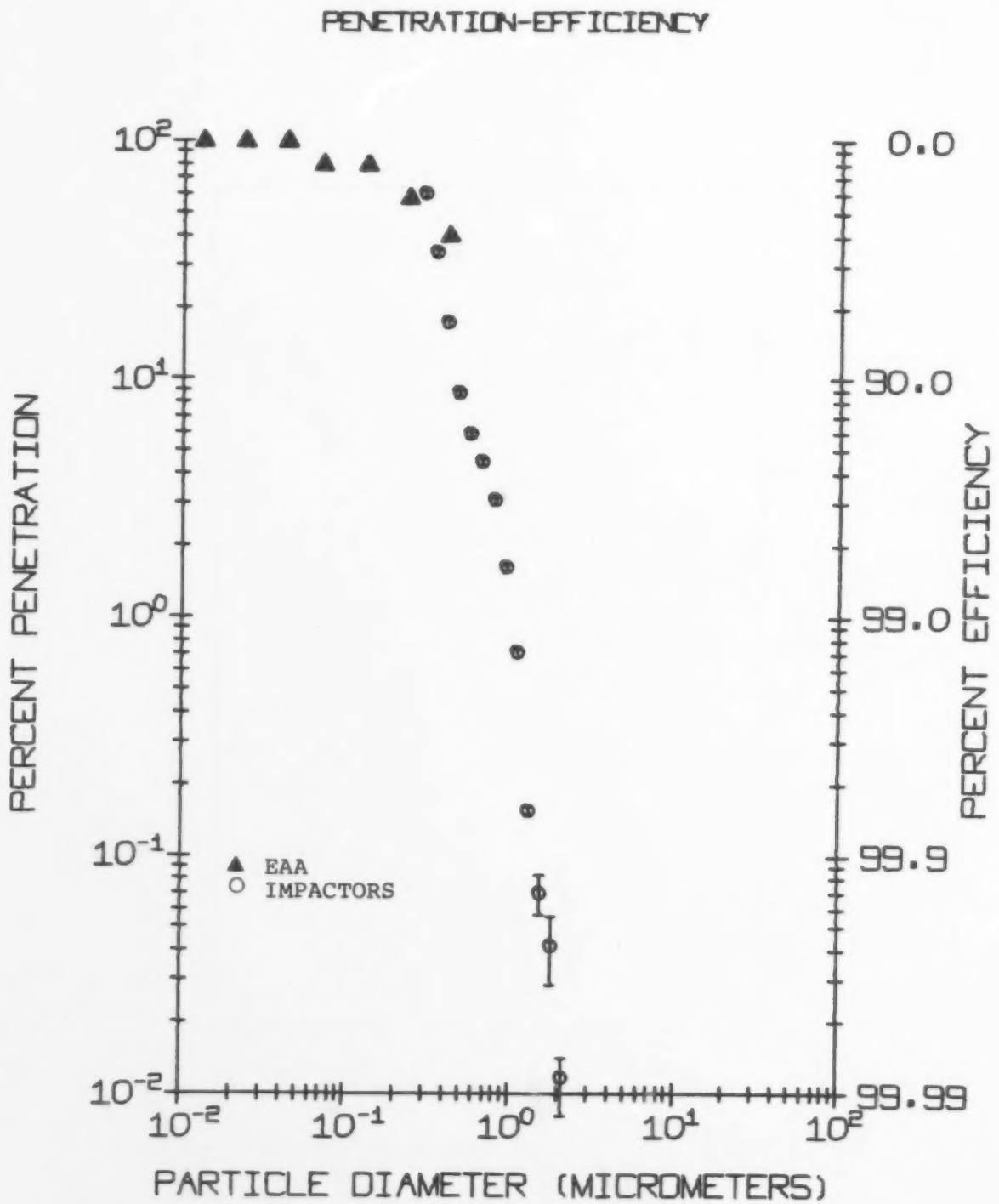

Figure 35. Fractional efficiencies based on electrical mobility and optical methods shown on a "physical" diameter basis. Also shown are fractional efficiencies from the cascade impactor data on a basis of Stoke's diameters.

on May 21, during which time the outlet concentrations were monitored with the EAA and the optical counter. No significant concentration changes were noted in the EAA data over this range of pressure drops; however, the optical counter data did show significant changes. In the 2 μm to 4 μm size interval, a 50% reduction in concentration was obtained by increasing the venturi pressure drop from 31 to 51 cm w.c. and a 35% reduction in concentration occurred in the 0.6 μm to 2.0 μm particle diameter range. These relative concentration changes are shown in Figure 36.

Table 19 summarizes the scrubber operating conditions throughout the test period. The liquid to gas ratio in the venturi portion of the scrubber during the tests was typically about 3.3 ℓ/DNCM.

SO_2 concentrations and collection efficiencies were also measured during the test program. Results of these measurements are given in Table 20, from which it can be seen that typically the SO_2 collection efficiency is about 80%.

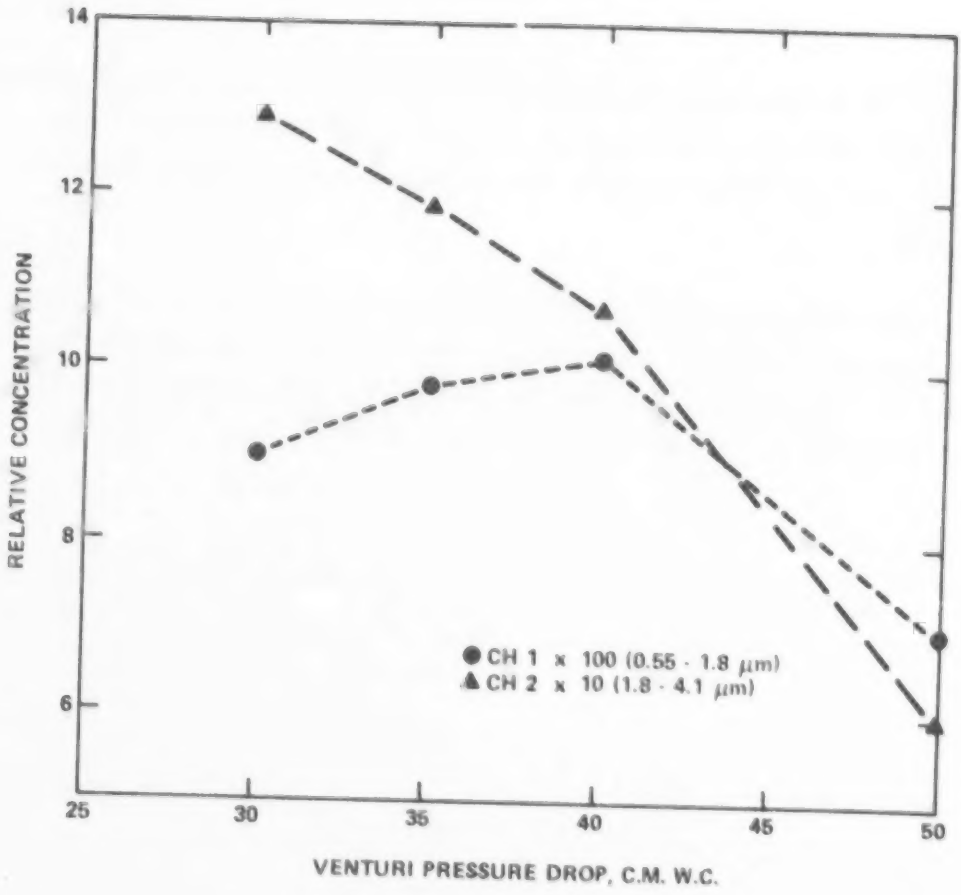

Figure 36. Relative outlet particulate concentrations in two size ranges as functions of venturi pressure drop.

Table 20

Colstrip Power Plant
Scrubber SO_2 Removal Efficiency

Date	Inlet SO_2 Concentration (ppm)	Reheater Outlet SO_2 Concentration (ppm)	SO_2 Removal Efficiency (%)
5-17-77	658	130	80.2
5-18-77	525	103	80.4
5-19-77	553	130	76.5
5-20-77	625	90	85.6

SECTION VIII

SUMMARY

The performance obtained by the devices described in this report can be compared with the expected performance figures for other types of scrubbers using the "cut diameter" method described by Calvert (1974) J APCA, 24:929. This method is based on the idea that the most significant single parameter to define both the difficulty of separating particles from gas and the performance of a scrubber is the particle diameter for which the collection efficiency is 0.5 (50%). Typical performance figures for the devices tested are shown in Figure 37, which is adapted from one presented by Calvert (op cit). It shows control device aerodynamic cut diameter graphed against power per unit flow rate (hp/1000 acfm). Also shown is the equivalent air pressure drop if all the power went into moving the volume of air through a flow resistance. The lines shown are theoretical and not experimental. Line 1 is for a sieve plate type scrubber with froth density F = 0.4 and hole diameter d_h = 0.3. Line 3 is for impingement plates and Line 4 is for a packed column from 1 to 3 m high and packing of nominal 2.5 cm diameter. Line 2a for venturi scrubbers (f = 0.25) represents the performance of venturi scrubbers in collecting hydrophobic particles while Line 2b (f = 0.5) represents the same class of scrubbers in collecting hydrophyllic particles.

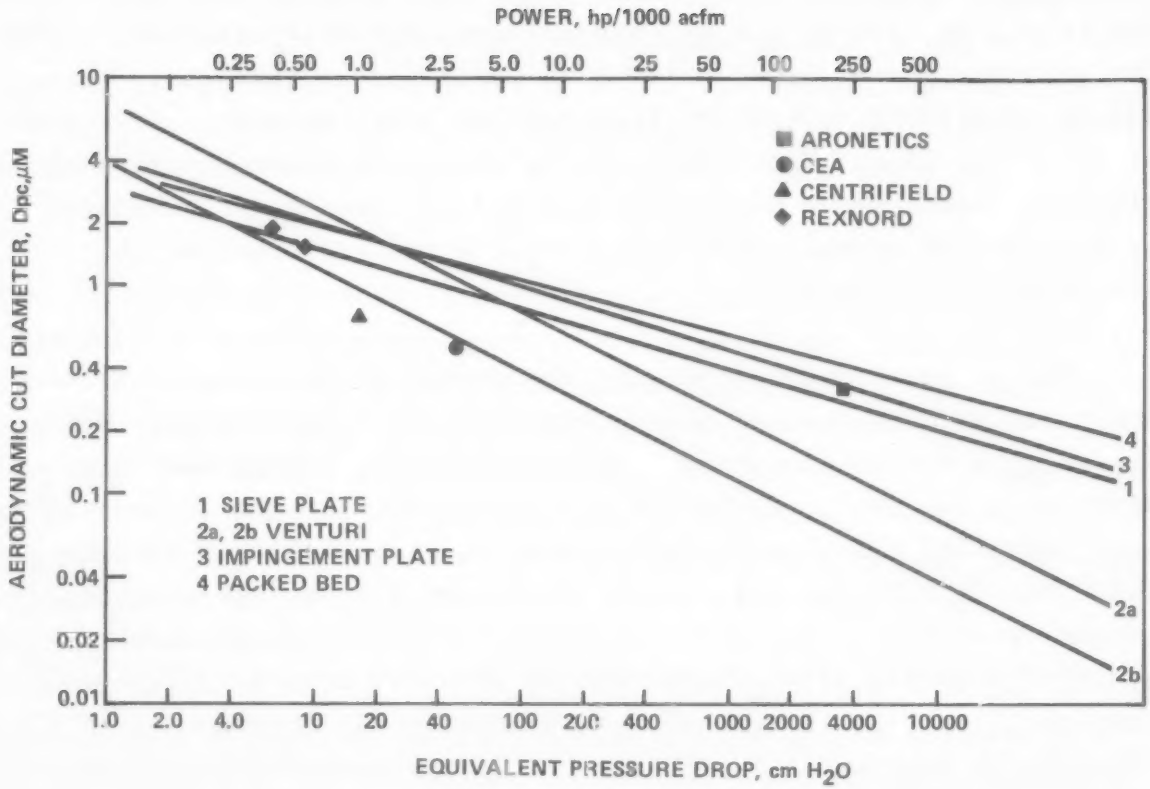

Figure 37. Representative cut diameters as a function of pressure drop for several scrubber types, after Calvert (1974), J. APCA 24:929.

The collection efficiency of the Aronetics Two Phase Scrubber determined by conventional total mass techniques on a source producing particulate having a mass mean diameter of about 3 μm was 95.1 and 96.7% for two days of testing. Measured fractional efficiencies were about 70% at 0.01 μm, about 35% at 0.05 μm, 35% at 0.1 μm, 99% at 0.5 μm, 99% at 1 μm, and 99.4% at 5 μm. The scrubber energy usage during the tests was approximately 633000 JOULES/SCM (17000 BTU/1000 SCF) at a net pressure rise of 12 1/2 to 16 in. H_2O. This energy usage was a result of using all the process waste heat available and may have been in excess of the minimum amount required to achieve the efficiencies obtained during these tests.

The collection efficiency of the Entoleter Inc., Centrifield Scrubber determined by conventional total particulate techniques on a source producing particulate having a mass mean diameter of about 100 μm was 99.50 and 99.73 for two days of testing. Measured fractional efficiencies were about 90% at 0.03 μm, about 50% at 0.05 μm, 20% at 0.1 μm, 50% at 0.5 μm, 90% at 1 μm, and 99.7% at 5 μm. The scrubber energy usage during the tests was approximately 2980 JOULES/SCM (80 BTU/1000 SCF) at a pressure drop of 17 cm (6.7 inches) w.c. The Centrifield Scrubber was not found to be substantially different in efficiency from the class of scrubbers known as conventional scrubbers. As shown in Figure 37, the power consumption for the Centrifield was somewhat lower but not substantially different from that for a well-designed venturi scrubber giving the same particulate collection efficiency.

The collection efficiency of the Rexnord gravel bed filter, determined by conventional (method 5) techniques on a source

producing particulate having a mass median diameter of about 200 μm ranged from 95% to 98% during three days of testing throughout which the collector was not operating in an optimum mode. Overall efficiencies determined from cascade impactor data during a second two-day test series were found to be 99.3% and 99.7%. The system pressure drop in the first test series ranged from 11.6 to 17.6 cm w.c. while during the second test series the system pressure drop ranged from 9.6 to 14 cm w.c. Measured fractional efficiencies were about 50% at 0.04 μm, zero or negative over the size interval from about 0.08 μm to 1.0 μm, approximately zero at 1 μm, 30% at 2 μm, and about 97.5% at 5 μm. The system energy usage during the tests was approximately 1780 JOULES/SCM (47.7 BTU/1000 SCF) at a pressure drop of 11.8 cm (4.7 inches) w.c.

Most of the devices tested to date under the novel device test program have been scrubbers. For this reason it has been convenient to compare their performance to a conventional venturi scrubber. The Rexnord gravel bed, while not a scrubber, has also been compared on the same basis as shown in Figure 37. It was determined that the power consumption of the Rexnord unit was somewhat higher but not substantially different from that of a well-designed venturi scrubber giving the same particulate collection efficiency.

The overall collection efficiency of the CEA variable throat venturi scrubber, determined by conventional (Method 17) techniques on a pulverized coal fired power boiler producing particulate having a mass median diameter of about 20 μm ranged from 99.12 to 99.50 during three days of testing. The venturi pressure drop ranged from 44.5 cm w.c. to 48.3 cm w.c. Measured fractional efficiencies were about 5% at 0.06 μm, 25% at 0.1 μm, 40% at 0.20 μm, 50% at 0.5 μm, 98.4% at 1.0 μm, and 99.99% at 2 μm. The system energy usage during the tests was approximately

7200 JOULES/DNCM (193 BTU/1000 SCF). SO_2 collection efficiency ranged from 76.5% to 85.6%.

A comparison of the performance of the CEA scrubber with that of conventional scrubbers of various types is shown in Figure 37.

The results of this comparison indicate that the scrubber tested performed somewhat better than a well designed conventional venturi scrubber operating at the same pressure drop; however, the difference may well be within the uncertainty in the measured performance.

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4. McCain, Joseph D., "Evaluation Of CEA Variable Throat Venturi Scrubber." SORI-EAS-77-650, Contract Number 68-02-1480, November 16, 1977.